Use of RDF to Dry Recovered Froth-Floated Glass

Final Report
October 1980

MASTER

Prepared for:

U.S. Department of Energy

Assistant Secretary for Conservation and Solar Energy

Office of Buildings and Community Systems

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ABSTRACT

An evaluation is made of the technical and economic feasibility of using waste- or refuse-derived fuel, at a resource recovery site, to provide the energy for drying glass recovered by froth flotation. A survey of equipment manufacturers and a technical study of an assembly, including a waste combusting unit, heat exchanger and dryer is conducted. An equipment package is defined and costed. The analysis shows that, at the present price of oil, and probably for the next decade, drying froth-floated glass is done more economically, simply and compactly with fuel oil than with refuse-derived fuel in a single-use, dedicated combustor.

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CHAPTER 1

OBJECTIVES, APPROACH AND RESULTS

OBJECTIVES

The purpose of this work is to evaluate the technical and economic feasibility of using waste or refuse-derived fuel, at a resource recovery site, to provide the energy for drying glass recovered by froth flotation.

The use of recovered cullet to manufacture new glass containers can save an estimated 0.15×10^{17} J/year (0.14×10^{15} Btu/year) in the U.S. (1).

In the economics of glass recovery, from mixed municipal waste, by froth flotation, the cost of fossil fuel to dry the product prior to shipping is in the neighborhood of \$2/ton (2), a significant cost when compared to the \$20 to \$30 which the glass might command on the market. Substitution of a lower (or even negative) cost waste fuel available at the recovery site might conserve fossil fuel and improve the economics, thus bringing about increased glass recovery across the country and indirect energy savings in the manufacture of new glass.

As an example, the output of recovered glass from the existing 650 tons per day plant at New Orleans, LA. (Recovery 1) is expected to be about 4 tons per hour. The initial and final moistures are estimated to be 12% and 0.5%, respectively. For practical values of the combustion and drying efficiencies, the energy input in the form of waste fuel is approximately 4.2 GJ/h

(4 M Btu/h). It would require 363 kg/h (800 lb/h) of a waste fuel of heating value 11.6 MJ/kg (5000 Btu/lb). This amount represents a small fraction, about 1%, of the total refuse-derived fuel (RDF) which could be produced on the site. On the other hand, savings effected by not having to landfill the material used as a waste fuel should be quite small.

On this basis, it was of interest to determine if, in order to save fossil fuel, a valid alternate solution might be to use waste fuel in a suitable combustor. The energy so raised would be used in the drying process, possibly through one or more heat exchangers. Validity was ascertained on both technical and economic grounds.

As a result of this study, conclusions on a recommended course of action are drawn, including the outline of an experimental program in the use of RDF to dry recovered glass.

APPROACH

In the process diagram sketched in Fig. 1-1, the main components of the equipment assembly are represented:

- (1) A waste fuel, or refuse-derived fuel, is used as the primary source of energy.
- (2) A combustor converts the fuel chemical energy into heat contained in the combustion gases.
- (3) A heat exchanger and combustion heat conveying medium transfers the heat from the combustor, into the dry, or alternatively into

Fig. 1-1. Process Components

- (4) A dryer and dryer heat conveying medium, which provides the heat input into the dryer. (Note: the media in (3) and (4) might be identical).

Each of the components was evaluated as regards input, output, efficiency, type, size, interfaces, price and how these factors affect the overall technical and economic solution.

The fuel was considered with respect to composition; combustion parameters; excess air; and composition, volume and temperatures of combustion gases. An analysis of a few examples of "available waste fuels" gives a range of expected characteristics.

The dryer type, materials and heat conveying medium are important in determining the interface with the combustor, as well as the overall size, complexity, efficiency and price of the assembly. Since froth-floated glass consists of particles in the 0.1 to 1.0 mm size range, special attention must be paid to dusting and entrainment of "fines" in those cases where the velocities of gases around the material to be dried are relatively large. Furthermore, to meet specifications, the dry material cannot contain more than a few parts per million of ceramics, and a sureproof way to avoid contamination is to prevent direct contact between combustion gases and their particulates with the material to be dried.

The combustor size is basically determined by the nature of the fuel and energy requirements of the dryer, account being taken of energy losses in intermediate equipment, such as a

waste heat boiler or heat exchanger.

The heat exchanger (if needed) would transfer heat from the combustion gases (or steam) to the dryer heat conveying medium.

As an example, Fig. 1-2 illustrates some of the possible options in this sequence of operations. Rotary - and fluidized bed dryers - are represented as possibly well suited to dry a fine, granular, sand-like product such as froth-floated glass, having a bulk density of approximately 1120 kg/m³ (70 lb/cft).

All of the above components were analyzed in part through a survey of equipment manufacturers and study of vendors' quotes and in part by assessing the importance of system parameters.

A particular assembly was selected which was thought to be of reasonable efficiency and cost, proven in practice under relatively similar conditions, and for which material losses and operational complexity appeared to be minimal.

These data were then used in computing alternative costs of this system vs. a "conventional one," using a combination oil heater-hot thermal fluid dryer.

RESULTS

Using realistic examples of waste fuels available at a recovery site, it was determined that for excess air on the order of 50%, exit temperatures of the combustion gases should be about 1800°F, in a modular ("starved air") incinerator.

The survey of dryers and drying methods conceivable for froth-floated glass, at a 4 TPH nominal throughput, showed no definite price advantage for any type of dryer. Dusting and fines

Fig. 1-2. Options in the Drying Process

carryover occurring in direct rotary or fluidized beds are of concern when dealing with a fine and polydisperse distribution of particles. This, and the contamination of the product by the combustion gases are avoided by using an indirect-type dryer and a thermal fluid as the heat conveying medium.

Several types of small-size combustors for RDF were examined. Modular incinerators quoted for the specified duty are more economical to purchase than the proposed fluidized bed combustors, and their ability to meet emission standards without elaborate fines recirculation and particulate collection devices is also advantageous.

Heating the thermal fluid in a high temperature flue gases heat-exchanger unit was chosen in preference to the use of a waste heat boiler as the least complicated of various options of same approximate cost, in view of the quoted high installation costs of steam dryers.

The economic analysis, based on present (end of 1978) prices of oil (fuel #2), shows that the solution of using RDF is economically disadvantageous compared to that of using fuel oil in a compact heater (thermal fluid unit). After accounting for the amount of the savings realized from the switch from oil to RDF (about \$20,000), a net yearly loss results which is of the same order of magnitude as the quoted amount. Correspondingly, the price of oil would have to quadruple over the next ten years to lead to a break-even situation, on the average.

An experimental procedure and layout for a possible technological experiment (disregarding the economic penalty) is

outlined and an instrumentation package for recording data on process throughput and efficiency is costed.

In conclusion, the analysis showed that drying recovered froth-floated glass, such as obtained at Recovery 1, New Orleans, is more economically, simply and compactly done with fuel oil than it could be with a single-use, dedicated combustor utilizing RDF as the energy source. However, investigations of the use of a larger combustion unit, using waste fuels, for a variety of on-site processes (glass drying, steam cleaning of metals, other steam uses), might well indicate that the approach makes technical and economic sense for multiple uses at the plant.

CHAPTER 2.

THE MATERIAL: FROTH-FLOATED GLASS

GENERAL

Froth-floated glass is a fine, granular, sand-like material recovered from municipal solid waste, more specifically its heavier fraction, by a succession of wet processing operations. These include jigging (3), sizing (4), and floating the material on the air bubbles of a froth (5,6). The wet glass product is then mechanically dewatered and dried (7).

As an example, at the 650 tons per day recovery plant in New Orleans, LA. (Recovery 1), the recovered product is dewatered by a vacuum filter, to a moisture content expected to be less than 10% (7). Final removal of the moisture (down to 0.5%, approximately) is presently carried out in cylindrical drum (Holoflite) dryer, using a thermal fluid at 343°C (650°F) for direct drying. Moisture is carried away by ambient air circulating at low speed, in the axial, upwards inclined direction towards the glass exit section and a venting stack. Draft is through natural circulation. The thermal fluid itself is heated in an oil heater, using #2 fuel, and capable of delivering 1.32 GJ/h (1.25 M Btu/h).

PROPERTIES OF THE GLASS PRODUCT FROM FROTH FLOTATION

Particle Size Distribution

Of paramount interest to the present study are the properties of the product to be dried. Recovered froth-floated glass was

studied experimentally at NCRR and results have been reported (5).

For later use in the calculations, a distribution of particle sizes by classes (as defined by U.S. Sieve Series mesh sizes) was obtained by averaging the results of two tests performed on a 3-stage cleaner product. The results are given in Table 2-A and Fig. 2-1.

TABLE 2-A: SIZE DISTRIBUTION OF CLEANED GLASS PRODUCT

Class (by mesh size, ASTM E 11-70)	(Arithmetic) Average Size of Particle in Class, $\bar{d}_i$ (mm)	Weight Percent in Class
120-170	.107	4
80-120	.151	4
60-80	.214	9
45-60	.302	20
35-45	.427	31
25-35	.604	23
18-25	.853	9

Tests NCRR 8327-1.0, 1117-1.0

As pointed out by Vanecek, "... for the characterization of particle sizes in a polydisperse mixture by a single quantity, variously defined mean particle diameters are employed. If the weight of the fraction of diameter d_i is M_i , [...]; then the weight mean diameter is

Fig. 2-1. Weight Fraction of Froth-Floated Glass Product by Size Classes

$$\bar{d} = \frac{\sum M_i d_i}{\sum M_i} \quad (2-1) \quad "$$

It is further stated that $\bar{d}$ from (1-1) is one of the recommended diameters "for calculations concerning fluidization." (9)

Using the data of Table 2-A, such a weight mean diameter, $\bar{d}$, can be computed from the individual class diameters d_i and weight fractions f_i , as

$$\bar{d} = \sum_i d_i f_i = 0.515 \text{ mm}$$

This dimension is in the (-30, +35) mesh range.

Other Properties

Other relevant properties of the glass product are listed, with the reference used as a source:

- . Material specific gravity (10): 2.6 (Range: 2.4-2.8)
- . Bulk density: - minimum: 960 kg/m³ (60 lb/cft)
 - maximum: 1440 kg/m³ (90 lb/cft)
 - nominal: 1120 kg/m³ (70 lb/cft)
 (The latter figure was determined experimentally prior to drying on a 55 gallon sample (11). It corresponds to a 57% void fraction, at zero moisture level.)
- . Thermal conductivity (12): 1.021 W/m, °C (0.244 cal/s, m, °C; 0.59 Btu/h, ft, °F)
- . Composition (13): in wt.%; SiO₂ = 72.5; Al₂O₃ = 2.5;
 CaO = 10.0; MgO = 0.5; Na₂O = 13.5;
 K₂O = 1.0

Specific heat, constant pressure (1 atm) (14):

$$C_p = 0.837 \text{ kJ/kg, } ^\circ\text{C} \text{ (0.20 kcal/kg, } ^\circ\text{C or} \\ \text{Btu/lb, } ^\circ\text{F) for "normal glass."}$$

COMMENTS

Recovered froth-floated glass is a sand-like, fine powdered material having a rather broad range of sizes, between 0.1 and 1.0 mm, approximately.

These characteristics will have a direct bearing on the technical choices to be made in the study. Wet froth-floated glass will tend to "cake," i.e., not be free-flowing, except at low levels of moisture. This will give a preference to drying devices in which the material is constantly kept agitated, both to prevent caking and to improve the overall heat transfer and vaporization of the moisture. Secondly, the relatively wide spectrum of particle diameter sizes should be expected to cause a significant elutriation of small particles at gas velocities sufficient to fluidize the large size fraction of the solids. Also, in direct drying by combustion gases, it is possible that the smaller particulate material would be entrained out of the dryer instantaneously, before any moisture has been removed.

*This value is in agreement with the one obtained from the above chemical composition, $C_p = 0.833 \text{ kJ/kg, } ^\circ\text{C}$ (0.199 kcal/kg, $^\circ\text{C}$ or Btu/lb, $^\circ\text{F}$).

CHAPTER 3

THE FUEL: REFUSE-DERIVED FUEL (RDF)

GENERAL

In this section, some examples of fuels that could be used as the source of energy in the glass-drying process are examined. To some degree, most fuels from waste require steps of preparation, going from the elementary (collecting screens oversize) to the more elaborate (densified light fraction).

The varieties of RDFs in existence, and the process flows to produce them in either "fluff" or "compacted" form (such as densified refuse-derived fuel, "d-RDF") are discussed elsewhere (2).

As shown in Fig. 1-2, the combustion gases from the RDF could be used to directly dry the glass, to raise steam, or to heat a thermal fluid. In order to design and size the various pieces of equipment (combustor, heat exchanger, dryer, etc.), it is necessary to derive values for some significant parameters, such as adiabatic and exit gas temperatures, gas composition and excess air.

FUEL AND COMBUSTION CHARACTERISTICS FOR TYPICAL WASTE FUELS

Properties and Composition

The first example fuel resulted from shredding of raw municipal solid waste, to <100 mm (4 inch), followed by flat

screening (6.4 mm or 1/4 inch), air classification and secondary shredding, to <25 mm (1 inch). The compositions and heating values were obtained from six samples, analyzed by an independent laboratory (15). The data are summarized in Table 3-A.

TABLE 3-A. COMPOSITION OF RDF (FIRST EXAMPLE)

Ultimate (Dry Basis)		Ultimate (As Received)		Proximate (As Received)	
C	0.4767	C	0.3603	Moisture	0.2442
H	0.0571	H	0.0703	Ash	0.1342
N	0.0082	N	0.0062	Volatile	0.5425
O*	0.2804	O**	0.4218	Fixed C†	<u>0.0791</u>
Ash	<u>0.1776</u>	S	0.0072	Total	1.0000
Total	1.0000	Ash	<u>0.1342</u>		
		Total	1.0000		

*By difference; includes S, Cl

**With hydrogen and oxygen present in 24.42% moisture

†By difference

Higher Heating Value: 13.51 MJ/kg (5,811 Btu/lb)

The composition of combustion gases (by weight and volume) resulting from the stoichiometric combustion of this fuel is given in Appendix A1.

It is found that the amount of standard air (1.3% moisture by weight, dry basis (15)) required for the stoichiometric combustion of one unit mass of fuel, w_a , is $w_a = 4.83$ (kg air/kg fuel).

Knowing the gas composition, the specific heat at 1 atmosphere, $\bar{c}_p$, is found to be, at zero excess air (E.A.) and on the average over the range of temperatures, $\bar{c}_p = 1.348 \text{ kJ/kg } ^\circ\text{C}$ (0.322 kcal/kg, $^\circ\text{C}$ or Btu/lb, $^\circ\text{F}$). Excess air E.A. is defined as the excess over 1 of the ratio of actual combustion air mass to the stoichiometric combustion air. After correcting for the amount of water produced per unit mass of fuel, the adiabatic combustion temperature, T^* , at zero excess air is $T^*_{\text{E.A.} = 0} = 1547^\circ\text{C}$ (2817 $^\circ\text{F}$).

If the amount of air used in the actual, non-stoichiometric combustion is w'_a (kg of air/kg of fuel), then $w'_a = w_a (1 + \text{E.A.})$. Appendix A1 gives details and results of calculations for the combustion of the above fuel and various values of the excess air. The adiabatic temperature $T^*_{\text{E.A.}}$ as a function of excess air E.A. is plotted in Fig. 3-1.

The small-scale combustion equipment likely to be used to generate the flue gases is expected to have a combustion efficiency (η_c) of about 0.8 reflecting radiative losses (which tend to be rather large for small units), leakage and drafts, unburned material, ash sensible heat, and other conductive and convective exchanges with the surroundings.

Accordingly, a flue gas temperature T_g is computed using the formula

$$T_g - 20 \approx 0.8 (T^*_{\text{E.A.}} - 20) \quad (^\circ\text{C}) \quad (3-1)$$

and is given as a function of E.A. in Fig. 3-1.

In order to avoid excessive exit temperatures of the combustion gases, and damage to the refractory walls, it is a rule of

Fig. 3-1. Adiabatic Combustion Temperature vs. Excess Air

good practice for small combustors' manufacturers to limit T_g to approximately 982°C (1800°F). For incinerators, a range of 982°C to 1204°C (1800°F to 2200°F) is expected (17). This 982°C (1800°F) temperature would require an excess air (obtained by extrapolation) of 0.46 or 146% of stoichiometric air, with an 80% efficient combustor, and the fuel under study. The mass of flue gas is 7.92 kg/kg of fuel. The composition of flue gases and other relevant quantities, for this value of E.A. (0.46), are given in Appendix A1.

If no further dilution occurs, the thermodynamic conditions of the flue gases at the combustor exit section are:

$$\text{Pressure: } P_{\text{gas}} \approx 101,320 \text{ Pa (14.7 psia)}$$

$$\text{Gas constant: } r_{\text{gas}} = 289.1 \text{ (J/kg, } ^\circ\text{K)}$$

$$\text{Temperature: } T_{\text{gas}} = 1257.1 \text{ } ^\circ\text{K (1803}^\circ\text{F)}$$

$$\text{Specific mass: } \rho_{\text{gas}} = 0.278 \text{ kg/m}^3 \text{ (0.0174 lb/ft)}$$

Prior to entering the dryer circuit, the gas temperature might have to be lowered to a temperature of about 649°C (1200°F). This is assumed to be done by adiabatic mixing of the flue gases with standard air at 20°C (68°F). The composition of the mixture is given in Appendix A1. The total quantity of diluted flue gas, w_a'' , is

$$w_a'' = 12.90 \text{ (kg flue gas/kg fuel)}$$

and the thermodynamic conditions of the diluted gas

$$P_{\text{gas}} \approx 101,320 \text{ Pa (14.7 psia)}$$

$$r_{\text{gas}} = 289.4 \text{ (J/kg, } ^\circ\text{K)}$$

$$T_{\text{gas}} = 922^\circ\text{K (1200}^\circ\text{F)}$$

$$\rho_{\text{gas}} = 0.380 \text{ kg/m}^3 \text{ (0.0237 lb/ft)}$$

These conditions should be considered typical of those prevailing at the inlet of the dryer, in the case of direct drying by the combustion gas.

Other potential feedstocks are readily available in resource recovery plants, in addition to the fuel discussed so far. These are:

Fuel "B"

This fuel fraction consists of the oversize fraction of raw waste (as-received) after a rotary screen (trommel) having large holes (>4 inch), and magnetic scalping. The composition of this fraction, as well as the screening efficiency, were measured at the NCRR Recovery 1 facility (18).

In the present calculation, an average between two runs reported in this reference (Runs #1 and #2) is taken to be typical. The efficiency of magnetic recovery is assumed to be 95%, believed to be realistic for a 2-stage, 90% efficient process (19). A breakdown in waste components is given in Appendix A1. The resulting properties are:

Higher heating value: $(HHV)_{B, \text{ dry}} = 18.82 \text{ MJ/kg}$
(8092 Btu/lb), dry basis

$(HHV)_B = 13.44 \text{ MJ/kg}$
(5778 Btu/lb), as received

Ash content of components was estimated (see Appendix A1), to yield an overall ash content, by weight:

$(Ash)_{B, \text{ dry}} = 9.2\%$, dry basis

$(Ash)_B = 6.6\%$, as received

Moisture: 28.6%

Fuel "C"

Fuel fraction "C" would be made of the "-4 inch" trommel undersize material, obtained as described above, from which magnetic materials have been scalped. After a step of air classification, the light fraction is retained as fuel "C".

Again, to determine the composition of the trommel undersize fraction, the average figures of Bernheisel, et. al. (18), runs #1 and #2, were adopted. Details are given in Appendix A1. It is found that "C" is a rather poor fuel, of higher heating value 9.50 MJ/kg. After additional screening to remove particulates under 1/4 inch, fuel "C'" is obtained and has the following properties:

Higher heating value:	$(HHV)_{C',dry}$	= 16.55 MJ/kg (7117 Btu/lb), dry basis
	$(HHV)_{C'}$	= 11.67 MJ/kg (5017 Btu/lb), as received
Ash content:	$(Ash)_{C',dry}$	= 15.4%, dry basis
	$(Ash)_{C'}$	= 10.9%, as received
Moisture:		29.5%

COMMENTS

It is seen that the RDF used as the first example above, and in the combustion parameters calculations, has gross properties very similar to fuels B and C'. It appears reasonable therefore to base the design on the RDF, the elemental composition of which is known, as a good technical approximation to fuels B or C'. Note that the use of fuel B rather than C' would result in lower charging rates, and possibly a more compact combustor, for a given desired Btu release per unit time.

CHAPTER 4

EVALUATION OF GLASS DRYERS

GENERAL

For the purpose of this analysis, it was essential to determine the geometrical, operational and economic characteristics of suitable dryers available on the market. More particularly:

- type of dryer (rotary, tube and shell, fluidized, etc.
- size
- efficiency of the drying process
- heat conveying medium required; flow rates; temperatures
- operational constraints (product size, dusting, carryover of fines, etc.)
- price; installation costs
- interface with the RDF combustor

Over a period of months, a survey of dryer manufacturers was conducted. Information was obtained on the phone, in writing or during technical visits. These data were analyzed and evaluated. This analysis resulted in recommendations for a preferred solution of the problem at hand.

In Appendix A2, some thermodynamic expressions and technical features of various types of applicable dryers are included for the reader's reference.

The following section examines the energy requirements of the drying process, using the terminology of Appendix A2.

ENERGY REQUIREMENTS FOR THE DRYING PROCESS

General

As discussed in Appendix A2, the energy total (per unit mass of wet feed) strictly required for the drying process to essentially zero moisture level comprises the following terms:

- a) Energy e_s required to heat the solids (glass particles) from ambient temperature t_a^* , to their final, or discharge temperature, $t_{s,f}$. If f_s is the weight fraction of the solids in the wet feed, and $c_{p,s}$ their specific heat at constant pressure,

$$e_s = c_{p,s} f_s (t_{s,f} - t_a^*) \quad (4-1)$$

- b) Energy e_l required to heat the water, in the liquid phase, from ambient temperature t_a^* , to the temperature, $t_{l,v}$, at which the liquid vaporizes. (Although some vaporization will occur at temperatures lower than 100°C (212°F), the conservative calculation should generally assume that $t_{l,v} = 100^\circ\text{C}$). If f_w is the weight fraction of the water in the wet feed, and $c_{p,w}$ the specific heat,

$$e_l = c_{p,w} f_w (t_{l,v} - t_a^*) \quad (4-2)$$

- c) e_{vap} is the energy required to vaporize the liquid at $t_{\ell,v}$ and atmospheric pressure. Given the latent heat of vaporization of the water, ℓ_{vap} ,

$$e_{\text{vap}} = f_w \ell_{\text{vap}} \quad (4-3)$$

- d) In the case of indirect drying: e_{air} is the energy transferred to the air carrying away the moisture after vaporization. If $\dot{m}_{1,\text{air}}$ is the mass flow of air per unit mass flow of wet feed; t_a^* , ambient temperature, and $t_{f,a}$, are the inlet and exit temperatures of the air, respectively; and $c_{p,a}$ is its specific heat

$$e_{\text{air}} = \dot{m}_{1,\text{air}} c_{p,a} (t_{f,a} - t_a^*) \quad (4-4)$$

Energy Input Required vs. Initial Moisture, Solids Discharge Temperature and Dryer Efficiency. Direct Drying.

In the case of direct drying, the combustion gases themselves will carry away the moisture, and e_{air} should not be included. Thus, the total energy, e_{tot} , for unit mass of wet feed, reads

$$e_{\text{tot}} = e_s + e_\ell + e_{\text{vap}} \quad (4-5)$$

in which e_s , e_ℓ , e_{vap} are computed from equations (4-1), (4-2), (4-3) respectively.

In Figures 4-1 and 4-2, the energy input per unit mass of wet feed, e_{in} , is defined as

Fig. 4-1. Energy Required Per Unit Mass of Wet Feed vs. Percent Feed Moisture Content (Solids Discharge Temperature = 107.2°C)

Fig. 4-2. Energy Required Per Unit Mass of Wet Feed vs. Percent Feed Moisture Content (Solids Discharge Temperature = 121.1°C)

$$e_{in} = \frac{e_{tot}}{\eta_d}$$

where η_d is the dryer efficiency defined in Appendix A2. e_{in} is plotted vs. f_w , initial moisture content (wet basis), a discharge temperature of the glass, $t_{s,f}$, equal to 107.2°C (228°F) or 121.1°C (250°F). Other constants are:

$$c_{p,s} = 0.837 \text{ kJ/kg, } ^\circ\text{C} \text{ (0.2 kcal/kg, } ^\circ\text{C; 0.2 Btu/lb, } ^\circ\text{F)}$$

$$t_a^* = 20^\circ\text{C (68}^\circ\text{F)}$$

$$t_\ell = 100^\circ\text{C (212}^\circ\text{F)}$$

$$l_v = 2.256 \text{ MJ/kg (539 kcal/kg; 970 Btu/lb)}$$

$$\eta_d = 1; 0.9; 0.8; 0.7; 0.6$$

Then:

$$E_{in} = \frac{1}{\eta_d} [0.837 (1 - f_w) (t_{s,f} - 20) + 4.187 f_w (100 - 20) + 2,256 f_w] \quad (4-6)$$

(kJ/kg)

This relation is tabulated in Table 4-A, for $t_{s,f} = 107.2^\circ\text{C}$ (225°F). When used in the following pages, interpolation is made linearly.

Energy Inputs Required: Indirect Drying

When the drying is indirect, a supplementary term needs to be added to e_{tot} in equation (4-5), namely

$$e_{air} = m_{l,air} (h_{f,a} - h_{20,a}) \quad (4-7)$$

where $h_{f,a}$, $h_{20,a}$ are the final and initial enthalpies of the air. The mass flow of air $m_{l,air}$ per unit mass flow of wet feed can be determined by specifying that in no case shall the relative humidity of the air and moisture carried out of the dryer exceed some fraction "k" of unity, at the temperature $t_{f,a}$ prevailing in the dryer exit section.

TABLE 4-A. ENERGY REQUIREMENTS FOR DRYING TO ZERO MOISTURE. (DIRECT CASE)

Energy Required/Unit Mass of Wet Feed: kJ/kg (Btu/lb)					
Wt. Percent Moisture (wet basis)	$\eta_d = 1.0$	$\eta_d = 0.9$	$\eta_d = 0.8$	$\eta_d = 0.7$	$\eta_d = 0.6$
0.20	577 (248)	640 (275)	721 (310)	823 (354)	961 (413)
0.15	451 (194)	500 (215)	563 (242)	644 (277)	751 (323)
0.12	374 (161)	416 (179)	470 (202)	535 (230)	626 (269)
0.10	326 (140)	361 (155)	407 (175)	465 (200)	542 (233)
0.75	263 (113)	291 (125)	328 (141)	374 (161)	437 (188)
0.05	200 (86)	221 (95)	249 (107)	284 (122)	333 (143)
0.	73 (31.4)	81 (35)	91 (39)	104 (45)	122 (52)

CONSTANTS

Amb. Temperature = 20°C (68°F)

Vaporization Temperature = 100°C (212°F)

Solids Discharge Temperature = 107.2°C (225°F)

$t_{f,a}$ is chosen relatively close to 100°C (212°F), to prevent recondensation at "colder" spots of the dryer exit; a margin of at least 20% to saturation at temperature $t_{f,a}$ should be kept for the same reason. For the design, we choose

$$\begin{aligned} t_{f,a} &= 90^\circ\text{C} \text{ (194}^\circ\text{F)} \\ k &= 0.8 = \frac{p_{\text{vapor}}}{p_{\text{saturation}}} \end{aligned} \quad (4-8)$$

For $t_{f,a} = 90^\circ\text{C}$ (194°F), $p_{\text{saturation}} = 0.692 p_{\text{atm}}$

The specific humidity, ω , i.e., the ratio of the mass of vapor to the mass of dry air in a unit volume of mixture is

$$\omega = \frac{m_{\text{vapor}}}{m_{\text{dry air}}} = \frac{\bar{M}_v}{\bar{M}_a} \frac{p_{\text{vapor}}}{p_{\text{air}}} = 1.118 \quad (4-9)$$

in which $\bar{M}_v$, $\bar{M}_a$ are the weight-average molecular masses of vapor and air, and p_{vapor} , p_{air} are the partial pressures of vapor and air.

The mass of vapor is computed from the sum of $m_{v,20}$, the mass of vapor present in the inlet air, assumed to be at 20°C (68°F) and saturated:

$$m_{v,20} = m_{\text{dry air}} \cdot 0.0147 \quad (4-10)$$

and the moisture released in the drying process, per unit mass of wet feed, $m_{v,g}$

$$m_{v,g} = f_w \quad (4-11)$$

Thus

$$\begin{aligned} m_{\text{vapor}} &= 1.118 m_{\text{dry air}} = m_{\text{dry air}} \\ &\quad 0.0147 + f_w \end{aligned} \quad (4-12)$$

The required mass of dry air is

$$m_{\text{dry air}} \times 1.103 = f_w \quad (4-13)$$

or the mass of humid air at 20°C (saturated) is

$$m_{\text{humid air}} = 0.920 f_w \quad (4-14)$$

The corresponding increase in enthalpy is

$$e_{\text{air}} = 34.66 f_w, \text{ kJ/kg (14.9 } f_w, \text{ Btu/lb)} \quad (4-15)$$

Going back to Table 4-A, it is seen that, compared to the total energy requirements (direct case), this term would only represent from 1.2% ($\eta_d = 1$) to 0.7% ($\eta_d = 0.6$) for an initial moisture level of 20%, and from 1.1% ($\eta_d = 1.0$) to 0.6% ($\eta_d = 0.6$) for an initial moisture level of 10%.

In view of the accuracy with which the dryer efficiency is known, this term could thus be disregarded with negligible loss of accuracy in the preliminary prediction of actual performance.

Using equation (4-15), the elements of Table 4-B were computed, similarly to Table 4-A above.

Application to an example: 59 Mg/h (65 ton/h*) recovery plant.

As an example, consider a plant processing 59.02 Mg/h (65 ton/h) of raw refuse over 10 effective hours, for an average throughput of 59.02 Mg/h (65 ton/h).

The rate of production of froth-floated glass, with a 75% recovery efficiency and an 8% content by weight of glass in the raw waste, is as a lower estimate (superscript "l")

*ton/h: short tons/hour

TABLE 4-B. ENERGY REQUIREMENTS FOR DRYING TO ZERO MOISTURE (INDIRECT CASE)

Wt. Percent Moisture (wet basis)	E_{in} , Energy Required/Unit Mass of Wet Feed: kJ/kg (Btu/lb)				
	if $\eta_d = 1.0$	if $\eta_d = 0.9$	if $\eta_d = 0.8$	if $\eta_d = 0.7$	if $\eta_d = 0.6$
0.20	584 (251)	649 (279)	730 (314)	835 (359)	972 (418)
0.15	456 (196)	507 (218)	570 (245)	651 (280)	761 (327)
0.12	379 (162)	421 (181)	472 (203)	542 (232)	630 (271)
0.10	328 (141)	365 (157)	412 (177)	470 (202)	549 (236)
0.075	265 (114)	295 (127)	333 (143)	379 (163)	442 (190)
0.05	202 (87)	223 (95)	251 (108)	288 (124)	337 (145)
0.	73 (31.4)	81 (35)	91 (39)	104 (45)	122 (52)

CONSTANTS

Amb. Temperature = 20°C (68°F)

Vaporization Temperature = 100°C (212°F)

Solids Discharge Temperature = 107.2°C (225°F)

Inlet Air Saturated: 20°C (68°F)

Exit Air: 80% R.H.; 90°C (194°F)

$$\dot{M}_{\text{glass}}^{(\ell)} = 0.06 \times 59.02 = 3.54 \text{ Mg/h (3.9 ton/h) (dry basis)}$$

Increases of up to 1 point in the glass content are likely, leading to an upper estimate (superscript "max")

$$\dot{M}_{\text{glass}}^{(\text{max})} \approx 4.0 \text{ Mg/h (4.4 ton/h) (dry basis)}$$

Based on the upper estimates, the net total energy required into the dryer (i.e., out of the combustion gases, steam, thermal fluid, etc.) is tabulated in Table 4-C, for the indirect case, representing the most conservative assumption (it can also be used for the direct case, as mentioned above).

As an illustration of costs (or possible savings) involved in the combustion of fossil fuels, consider the case where oil is combusted, with an efficiency of combustion:

$$\eta_c \approx 0.85$$

and at a fuel cost of 10.57 ¢/liter (40¢/gallon), and HHV = 39.86 MJ/liter (1.43×10^5 Btu/gallon), or

$$\text{\$2.65/GJ (\$2.80/M Btu)}$$

The elements of Table 4-D are expressed in yearly costs (250 days year), for glass drying.

It is seen that, if the wet glass, coming out of a vacuum filter, is at a (relatively high) 12% level of moisture, and for a range of dryer efficiencies from 80% to 60%, the maximum oil savings to be expected by substituting a free fuel to oil are, on a yearly basis,

TABLE 4-C. ENERGY REQUIRED FOR DRYING GLASS (590 Mg/day) PLANT

Total Energy Required/Unit Mass of Wet Feed: GJ/h ($\bar{M}$ Btu/h)					
Wt. Percent Moisture (wet basis)	$\eta_d = 1.0$	$\eta_d = 0.9$	$\eta_d = 0.8$	$\eta_d = 0.7$	$\eta_d = 0.6$
0.20	2.33 (2.21)	2.60 (2.46)	2.91 (2.76)	3.33 (3.16)	3.88 (3.68)
0.15	1.81 (1.72)	2.03 (1.92)	2.28 (2.16)	2.60 (2.46)	3.04 (2.88)
0.12	1.51 (1.43)	1.68 (1.59)	1.89 (1.79)	2.16 (2.05)	2.51 (2.38)
0.10	1.31 (1.24)	1.46 (1.38)	1.65 (1.56)	1.88 (1.78)	2.19 (2.08)
0.075	1.06 (1.00)	1.18 (1.12)	1.33 (1.26)	1.51 (1.43)	1.76 (1.67)
0.05	0.81 (0.77)	0.89 (0.84)	1.00 (0.95)	1.15 (1.09)	1.35 (1.28)
0.	0.30 (0.28)	0.32 (0.31)	0.36 (0.34)	0.42 (0.40)	0.48 (0.46)

590 Mg/day (650 t/day) Plant; 3.98 Mg/h (4.4 t/h) Glass, Dry Basis

CONSTANTS:

See Tables 4-A, 4-B

TABLE 4-D. YEARLY (OIL) FUEL COSTS FOR DRYING GLASS
(590 Mg/day Plant)

Fuel Costs, Thousands Dollars					
Wt. Percent Moisture (wet basis)	$\eta_d = 1.0$	$\eta_d = 0.9$	$\eta_d = 0.8$	$\eta_d = 0.7$	$\eta_d = 0.6$
0.20	18.18	20.24	22.71	26.00	30.28
0.15	14.15	15.80	17.77	20.24	23.69
0.12	11.76	13.08	14.73	16.87	19.58
0.10	10.20	11.35	12.83	14.64	17.11
0.075	8.23	9.21	10.37	11.76	13.74
0.05	6.33	6.91	7.82	8.97	10.53
0.	2.30	2.56	2.88	3.29	3.84

CONSTANTS:

Oil at 10.57¢/liter (40¢/gallon)

Oil Heater 85% Efficient

See Tables 4-A, 4-B, 4-C

from \$14,730 to \$19,580 if oil costs 10.57¢ a liter
(40¢ a gallon)

from \$29,460 to \$39,160 if oil costs 21.14¢ a liter
(80¢ a gallon)

from \$44,190 to \$58,740 if oil costs 31.71¢ a liter
(\$1.20 a gallon)

These figures, and particularly the ones corresponding to the present price of oil (end of 1978: 10.57¢/liter or 40¢/gallon), should be kept in mind when evaluating in the following sections, the costing of the equipment needs for: the combustion of RDF; the transfer of heat to the heat conveying medium, and the drying of the glass itself.

DIRECT ROTARY DRYERS

General

As discussed in Chapter 2, froth-floated glass is a relatively fine, sand-like material, with a particle size distribution in the 20 x 140 mesh range (.84 to .106 mm). Like sand, it is free flowing only when dry, and otherwise tends to form a dry superficial crust or "cake." For materials having these characteristics, dryer practice recommends equipment belonging to one of three types (see Appendix A2):

- direct, rotary
- indirect, rotary
- fluidized bed

Direct Rotary Dryers for Froth-Floated Glass

Geometry and design characteristics are discussed in Appendix A2.

In this particular case (drying froth-floated glass), the co-current configuration should be preferred, to avoid the risk of exposing the already dry product (near its exit section) to hot gases at a temperature (1200°F or more) which might exceed the softening temperature of the glass (1100°F, approximately) (19).

One problem of particular concern is the carryover of fine particulates, in the wide size range shown in Fig. 2-1.

Direct rotary drying requires combustion gas mass flow rates which are relatively high, in the 1.02 to 1.27 m/s (200 to 250 FPM) range, in order to:

- promote turbulent heat transfer between the gases and the solids
- insure a sufficient heat delivery rate into the dryer, for given specific heat of the gases, temperature drop along the dryer and temperature at the inlet, according to

$$\begin{aligned}\text{heat delivered} &= m_{\text{gas}} (t_{\text{gas, in}} - t_{\text{gas, out}}) c_{p, \text{gas}} \\ &= \rho_{\text{gas}} A_{\text{gas}} V_{\text{gas}} (t_{\text{gas, in}} - t_{\text{gas, out}}) c_{p, \text{gas}}\end{aligned}$$

in which A , ρ , V refer to the inlet section, and gas specific mass and velocity at the inlet.

It means that, at 1.02 m/s (200 FPM), a typical dryer, 1.83 m (6 ft) in diameter, D , and 7.32 m (24 ft) in length, L ($L/D = 4$), is traversed by the gas stream in 7.2 seconds. A

particle of sufficiently small diameter could be carried away with the gases and out of the dryer in a time which might be only a very small fraction of the design residence time, t_r , in the dryer.

The allowable mass flow rate per unit area of free section of the dryer, $\dot{m}_{\text{gas}}/A_{\text{gas}}$, will depend on the nature of the solids, the inside geometry of the dryer and the amount of carryover considered tolerable. Dusting and carryovers are said to be more severe for co-current designs, since at the exit, there is no wet feed playing the role of an impinger as in counter-current geometries.

It is often pointed out that final design should be "frozen" only after a series of tests on a "new" material to be dried, as is the case here for froth-floated glass. In fact, all quotes received from rotary dryer manufacturers were given, contingent upon tests run with the actual feedstock.

For the present technical assessment, we proceeded to evaluate likely values for the carryover of glass particles of various sizes. The developments and results are given in Appendix A3. The main result is that excessive entrainment of fines, in the approximate range <0.4 mm, or affecting as much as 49% by weight of the product, is to be expected at those velocities (or gas mass flow rates per unit area of free section of the dryer) which are required to dry the material, and for technically reasonable values of the dryer diameter.

To which extent this might be alleviated by the shape and location of breechings, or by the fact that collected fines might

have been dried during transport and collection, is only a matter of conjecture, which would require extensive test data.

It should be emphasized that the manufacturers explicitly make allowance for a (largely unknown) amount of dust and carry-over by providing the exhausted gas a cyclone, baghouse or other inertial collection device. This equipment, to which a forced draft (for the gases) and an induced draft (for the clean cold gas) fan have to be added, will obviously significantly add to the cost and complexity of the assembly.

For lack of experimental confirmation that the fines carry-over problem can be solved satisfactorily and simply, the choice of a direct rotary dryer entails a definite risk of insufficient drying time for the lower-size fraction of the product.

Summary of technical and cost data

The data gathered from quotes obtained from various manufacturers are summarized below;

Length: 4.9 to 9.2 m (16 to 30 ft)

Diameter: 0.92 to 1.83 m (3 to 6 ft)

L/D: 3.1 to 6.7

Drying

Efficiency: 55 to 75%

Cost Range: \$28,000* (one quote); \$55,000-\$75,000 (three quotes); \$95,000 (one quote)

INDIRECT DRYERS

For a general discussion of the features of indirect dryers, the reader should consult Appendix A2.

*Very few design data available on this quote.

Steam-heated Dryers

Quotes were solicited from six manufacturers and quotes received from three. Among the quotes received, two were for tube and shell steam-heated dryers, with the steam inside the tubes; the third one had the steam outside the tubes, and the glass and draft air inside these.

The following give approximate ranges for parameters of interest, as determined from these quotes.

Length: 4.6 to 10.7 m (15 to 35 ft)

Diameter: 1.22 to 1.83 m (4.0 to 6 ft)

L/D: 4 to 6

Drying

Efficiency: 55 to 85%

Cost range: \$65,000 to \$95,000

Indirect Dryers: Thermal Fluid or Molten Salts

This type of dryers, described in some detail in Appendix A2, was evaluated on the basis of three quotes from manufacturers and available technical literature.

Typical values of the parameters for thermal fluid or molten salts indirect dryers are given below for a 5 Mg/h glass maximum throughput of glass having an initial moisture of 12%.

Heat conveying medium temperature: 315°C to 371°C

(600°F to 700°F)

Diameter (rotor) or width of

trough*: 1.2 m (4 ft)

*Two 2-ft diameter screws

Length of rotor: 3.1 m to 3.7 m (10 to 12 ft)

L/D: 2.5 to 3

Drying efficiency: 70 to 75%

Cost range (including peripherals**) \$90,000 to \$115,000

FLUIDIZED BED DRYERS

Appendix A2 discusses this type of equipment and the elutriation of the smaller particulates from the dryer bed. It is estimated that approximately 37% (the population of the first four classes in Fig. 2-1) of the material, by weight, might be carried away from the bed.

If insufficiently dried, these particles would have to be recycled, resulting in a high circulating load and high dust loading and abrasion in the collection cyclone.

Manufacturers of fluidized bed dryers were generally reluctant to quote or dimension their units without testing this unfamiliar material, at a modest fee.

Data on the quotes received are summarized below.

Sizes: 1.2 m to 1.7 m (4 ft to 5.5 ft) diameter,

3.7 to 4.6 m (12 to 15 ft) high

Gas temperature, unit: 649°C (1200°F)

Drying efficiency: 60%

Cost range: \$90,000 (one continuous unit) to \$136,000

(two 227 kg (500 lb) capacity units, batch process)

**Such as baghouse and exhaust fan

Fluidized bed dryers did not appear to have any significant price advantage over other types of dryers. The need to collect (and possibly recycle) as elutriated fines as much as 37% of the load (a polydisperse suspension) would be a significant drawback -- so would the mixing of the materials with particles of fly ash if combustion gases are used directly as the heat conveying medium. Finally, some advantageously-priced units were thought to be semi-prototypes, not off-the-shelf hardware.

THROUGH-CIRCULATION DRYERS

Although this type of dryer, described in Appendix A2, might permit the use of combustion gases, it has the inconvenience of requiring a particulate collection system downstream of the discharge section and it will result in an intimate mixing, with possible contamination, of the recovered froth-floated gas.

These units are relatively more expensive than indirect (thermal fluid) or steam-heated units. A slight price advantage (\$10,000 or so) seems to exist compared to an indirect dryer/heat exchanger combination.

Typical data are given below.

Length: 4.9 m (16 ft)

Diameter: 1.6 m (5.2 ft)

L/D: 3.1

Drying efficiency: 71%*

Estimated price: \$95,000

*80% efficiency of combustion is assumed

CONCLUSIONS

This review of dryers characteristics and prices indicates that a number of designs are technically and economically comparable to dry froth-floated recovered glass.

Elutriation of fines or particle entrainment would be a matter of concern with fluidized bed or direct rotary dryers. This is due to the fact that froth-floated glass is a relatively fine material, in size ranges varying by one order of magnitude (from 0.1 to 1.0 mm), and that considerable volumes of gas are required to directly dry the wet material. Furthermore, for this type as for the through-circulation dryer, there would be intimate mixing of the particulates contained in the combustion gases and the recovered glass product.

These disadvantages, and the need for a particulate collection system, forced or induced draft blower, or feedwater conditioning system do not exist in the case of indirect dryers heated by a hot fluid. Their efficiency is high (in the low to mid-seventies), their complexity minimal, and their price advantageous, making them the most suitable candidates for the required duty.

CHAPTER 5

RDF COMBUSTOR

GENERAL

The present section examines the sizing and design parameters of the combustor using RDF as a waste fuel to generate the input heat into the drying process.

The fuel has been discussed in some detail in Chapter 3. It was seen that among the variety of fuel fractions obtainable from either trommel oversize or undersize material (possibly after a step of air classification), typical values might be those of fuel "B".

Moisture content: 28% (wet basis)

Higher heating value: 13.43 kJ/kg (5778 Btu/lb) (as-received)

Ash content $\approx$ 7% (as-received)

Bulk density: 80 kg/m³ (5 lb/cft)

The breakdown by components was given in Appendix A1, and was specified to manufacturers of combustion equipment from whom quotes were solicited.

At the low consumption rates of this fuel needed to dry the recovered glass, 910 kg (10³ lb) maximum, multi-chamber, refractory-lined incinerators are normally suitable in the vertical configuration. However, at furnace temperatures of 760°C to 982°C (1400°F to 1800°F), the excess air required is in the 100%

to 200% range. Meeting emission standards can be complicated and costly. Although they may be equipped with low pressure heat recovery boilers, the efficiency is relatively low. The requirement for auxiliary fuel support might be as much as 15% of the total Btu output.

The starved air incinerator includes a primary chamber where the solid waste is incompletely combusted by an amount of controlled air below the stoichiometric level. Combustion is completed in a secondary mixing chamber where additional combustion air is injected or induced. Off gases and particulates are efficiently burned at a temperature of 982°C to 1093°C (1800°F to 2000°F), with only minimum (and in some cases negligible) energy input from an auxiliary burner in the secondary chambers. When properly designed and controlled, these units (also referred to as "modular" incinerators) will, according to the manufacturers, meet Federal particulates emission standards.

A third type of combustor is the fluidized bed furnace. Here the fuel is injected and burned into a suspended bed of particles of mineral matter, such as ash, limestone, olivine, etc. The bed material is used to promote efficient, complete combustion and might play a role in removing pollutants (such as sulfur). Pollution control on the combustion products might be simpler. Typical bed temperatures would be no less than 871°C (1600°F), for odor destruction, and generally no more than 982°C (1800°F), to prevent fusion of the ash particles. A problem with the fuels considered here is that sizing of the feed material is non-uniform, which might result in poor combustion of

the larger particles or excessive fly ash entrainment. The entrained particles in the hot combustion gases would cause tube fouling in a waste heat boiler if the number of passes exceeded 2. Particulates are removed in a collection device, generally a hot cyclone. Another approach is to heat the steam (or thermal fluid) tubes in the bed itself for direct heat transfer without a waste heat boiler.

After a sizing of the combustor by feedrate of the "typical fuel," manufacturers proposed equipment and quotes will be reviewed.

COMBUSTOR: INPUT FEEDRATE

As was discussed, a nominal duty for the selected glass dryer would be:

Feedrate (wet feed): 4 Mg/h (8,810 lb/h)

Initial moisture: 12%

Residual moisture: negligible

Duty: 10 hours a day

From Table 4-B, for a drying efficiency of 0.6 and 0.8, respectively, it is seen that the input energy into the dryer is:

for $\eta_d = 0.6$, 630 kJ/kg feed (271 Btu/lb) or

2.52 GJ/h (2.39 M Btu/h)

for $\eta_d = 0.8$, 472 kJ/kg feed (203 Btu/lb) or

1.89 GJ/h (1.79 M Btu/h)

Average and Conservative Gross Energy Input Requirement Into the Combustor

The survey of manufacturers (Chapter 4) has shown that a realistic average value of the drying efficiency, for indirect

dryers of the thermal fluid, molten salt or steam type, is 0.7. For this "average" case, the energy required into the dryer is:

"AVERAGE" CASE: 542 kJ/kg (233 Btu/lb)

2.16 GJ/h (2.05 M Btu/h)

A conservative estimate is obtained by combining an expected efficiency of 0.6 with the maximum expected feedrate of 5 Mg/h (11,013 lb/h):

"CONSERVATIVE" CASE: 630 kJ/kg (271 Btu/lb)

3.14 GJ/h (2.98 M Btu/h)

From manufacturers' figures, the combustion efficiency, accounting for heat losses, unburned material and the use of the higher heating value, is about 0.8. Thus, of the energy originally present (as higher heating value), in the input waste, approximately 80% is available as enthalpy in the combustion gases, at the combustor exit section.

The waste heat boiler or heat exchanger (combustion gases to thermal fluid) recovers from 65% to 75% of this heat content, as the input into the dryer.

The ratio of the input energy into the dryer to the fuel input energy, η_* , ranges from 0.52 to 0.6.

For a practical value of this ratio, $\eta_* = 0.53$ (see Chapter 6), the input feed requirements, with a typical refuse-derived fuel, such as fuel "B" (Appendix A1), of higher heating value 13.44 MJ/kg (5778 Btu/lb) are, in the two cases defined above

$\eta_* = 0.53$, "AVERAGE" CASE: 4.07 GJ/h (3.86 M Btu/h)

304 kg/h (669 lb/h) of fuel

$\eta_* = 0.53$, "CONSERVATIVE" CASE: 5.92 GJ/h (5.61 M Btu/h)

441 kg/h (972 lb/h) of fuel

In any case, it is seen that the waste fuel input rate into the combustor would not exceed 454 kg/h (1,000 lb/h).

In what follows, the "design case" is assumed to correspond to the average conditions, i.e., an indirect dryer, of efficiency 0.7, heated by a combustor/boiler or combustor/heat exchanger assembly, which itself has a combined thermal efficiency, η_* , of 0.53. In this case, 304 kg/h (669 lb/h) of waste fuel are required to process 4 Mg/h (8810 lb/h) of wet feed at 12% initial moisture.

STARVED AIR INCINERATORS: POSSIBLE EQUIPMENT

Quotes and technical proposals were received from three vendors of starved air (or modular incinerators).

The equipment supplies would include:

- the incinerator unit
- a semi-automatic (ram-type) waste fuel loader
- the temperature control system (in secondary chamber)
- a loader water spray

These items make up the combustor package itself. If the combustion gases cannot be directly used (after dilution with cold air), the energy is recovered either in the form of steam in a waste heat boiler or hot oil in a heat exchanger (energy recovery) unit packaged as a separate unit with its own structure.

Erection, installation, start-up and training can be procured at extra cost.

From the quotes, it was readily apparent that, for a given energy input into the dryer, the size of these units (even without heat recovery systems) is fairly impressive. For example: a unit 7.3 m (24 ft) long, 3.7 m (12 ft) wide, 8.2 m (27 ft) high, weighing 22,700 kg (50,000 lb) provides a 2.11 GJ/h (2 M Btu/h) net input into the dryer.

A summary of data for this type of combustor is given below.

Modular Incinerators

Size (approx.): 7.3 m (24 ft) length; 3.7 m (12 ft) width; 8.2 m (27 ft) height

Weight (approx.): 22,700 kg (50,000 lb)

Input feedrate (for 2.11 GJ/h (2 M Btu/h) output):

295 to 318 kg/h (650 to 700 lb/h) of waste

Consumption of auxiliary fuel: negligible (except for startup and shutdown)

Temperature of combustion gases: 982°C (1800°F)

Price range (installed, without heat recovery)*: \$90,000 to \$110,000

Quoted efficiency: input feed to steam: 55 to 65%

input feed to thermal fluid: 50 to 55%

FLUIDIZED BED COMBUSTORS: PROPOSED EQUIPMENT

With the same specifications for the fuel and drying objective (or net heat input into the dryer), quotes were solicited

*Best estimate for waste heat boiler was deducted from package incinerator-boiler price when vendor declined to itemize.

for fluidized bed-type combustors using RDF as the fuel. Of the vendors contacted, three replied with documented technical offers and price quotes.

Since there are wider variations in design between fluidized bed furnaces than between starved air incinerators, a brief discussion of proposed designs and types might be in order.

Four quotes (one verbal) were received for fluidized bed combustors deemed to be suitable for the combustion of a RDF such as described above (fuel "B").

One of the proposed designs has a rectangular cross-section. Ambient or preheated air is introduced in the lower part by distribution orifices to fluidize the bed. The fuel can be fed by a variety of devices (vibratory, screw, conveyors) depending on the application and type of fuel. Start-up burners located above the bed and fueled by natural gas or oil are used to bring the bed to temperature after a shutdown. The media material is selected for each particular case, sand or possibly another material to effect chemical capture of the sulfur oxides, when required. In the overbed zone, combustion of the entrained combustible particulates occurs. Quenching by air also takes place in this zone, if desired. The outlet duct is horizontal, at the top side of the furnace, to minimize ducting and expansion problems. The unit is rated to deliver 1816 kg/h (4000 lb/h) of 10.2 atm. (150 psi) saturated steam.

Another design would immerse the tubes containing the thermal fluid in the bed itself, and a heat exchanger would also be used to preheat the fluidizing air with waste heat from the combustion

unit. Heat recovery is in the form of hot thermal fluid.

The third fluidized bed burner is cylindrical. It contains inert particles (olivine) as the bed media. Injection of the fuel material is from the top and combustion takes place in the vapor space (above the bed) as well as in the bed itself. The high temperature gases, quenched by overfire air, are controlled at 649°C (1200°F), approximately, at the inlet of the heat recovery boiler. After two passes in the horizontal fire-tube boiler, the cooled off-gases are cleaned in a particulate collector (cyclone type) and are induced to stack by an I.D. fan. Heat recovery is in the form of steam.

The last (verbal) quote was for a combination fluidized bed combustor/fluidized bed dryer. The fluidized bed dryer was 1.2 m (4 ft) in diameter, 3.7 to 4.6 m (12 to 15 ft) in height. The off-gases from the combustor, cooled by dilution with ambient air, enter the fluidized bed dryer at 649°C (1200°F). This combination was still experimental, and not an off-the-shelf item. Not surprisingly, to confirm the validity of the concept using RDF as a fuel to dry glass, a test at nominal cost (a few thousand dollars) was strongly suggested by the manufacturer. Hard data obtained on these unfamiliar materials within the still largely empirical fluidized bed environment might indeed prove very valuable to project engineers and system designers.

The significant parameters of these fluidized bed combustors are summarized below:

Diameter (or equivalent diameter): 0.9 to 1.2 m (3 to 4 ft)

Height: 6.1 m (20 ft)

Maximum energy/h (delivered to steam or thermal fluid): 2.22 GJ/h (2.1 M Btu/h)

Efficiency (input feed to thermal fluid or steam): approx. 75%

Feedrate: 220 kg/h (485 lb/h)

Assembly outside dimensions: 7.9 m (26 ft) length; 10.7 m (35 ft) high; 4.6 m (15 ft) width

Price (not installed): \$110,000 to \$120,000

Cost of shipping and installation*: \$30,000 to \$40,000

RECOMMENDED CHOICE

From the above data, a comparison between the starved air incinerators and fluidized bed combustors shows that either approach can be used for refuse-derived fuel (such as fuel "B") to raise steam or heat a thermal fluid.

The fluidized bed combustor has higher announced efficiency (75% vs. 55%) in the conversion of input feed energy into steam or thermal fluid. This efficiency might be on the optimistic side (except in the case where the thermal fluid tubes are immersed in the bed). Assuming an efficiency of 75% for the waste heat boiler alone (used in a combined 60% efficiency for starved air combustors), a combustion efficiency of 90% in the fluidized bed would result in a total efficiency of 68%. In any event, somewhat lower use of the waste fuel is only of marginal interest, since the fuel is free or has a negative value.

*including freight and installation of heat exchange equipment or waste heat boiler

The size and floor space for these units are comparable. Attrition of bed material and fine elutriation of particles have to be controlled by the pollution control equipment, in the fluidized bed case, whereas modular starved incinerators are installed with integral off-gases cleanup equipment.

Modular incinerators are an off-the-shelf design in burning a variety of municipal, manufacturing or commercial wastes. In the municipal environment, units including energy recovery in the form of steam are presently operating (20). There is no similar depth or breadth of experience in fluidized bed combustors using RDF, although, the approach has proved successful in a limited number of experimental tests.

The modular incinerators appeared, from this survey, to have a slight but definite price advantage on the fluidized bed combustors.

On this basis, for reasons of wider industrial practice, operating experience, clean exhaust gases without extensive additional pollution controls and price, it is recommended that a modular incinerator be used, assembled in combination with a heat exchanger (for the thermal fluid) or waste heat boiler (for steam).

The sizing and costing of the heat exchanger or heat recovery boiler is examined in the next chapter.

CHAPTER 6

HEAT EXCHANGER ASSEMBLY

GENERAL

If an oil-fired heater is used for heating the thermal fluid of the indirect dryer, heat is transferred efficiently (85%) in a compact unit.

If an RDF combustor is used to generate the heat, combustion gases exist from the modular unit at approximately 982°C (1800°F), possibly diluted with tertiary air to a temperature of 649°C (1200°F), to be compatible with the materials of the heat exchanger. The hot combustion gases flow through a cross-flow arrangement of tubes containing the thermal fluid (Fig. 6-1). The rows which are further downstream and in contact with cooler gases can be outfitted with annular rings (or fins) to increase the heat transfer area, for given outside dimensions and pressure drop.

Quotes were solicited from 10 manufacturers of air/gases to liquid heat exchangers, and two detailed quotes received are reviewed below.

SIZING OF HEAT EXCHANGER UNIT

Consider a cross-flow, one-pass heat exchanger in which the hot combustion gases flow over staggered rows of tubes in which the thermal fluid circulates. Since the enthalpy content of the heat conveying medium is proportional to $t_{in} - t_{ref}$, with t_{in}

Fig. 6-1. Cross-flow Heat Exchanger

the gross inlet temperature and $t_{ref} = 20^{\circ}\text{C}$ (68°F), the drop in gas temperature required for a 65% efficiency of the heat exchanger is

for $t_{in} = 482^{\circ}\text{C}$ (1800°F): from 482°C to 357°C
(1800°F to 674°F)

for $t_{in} = 649^{\circ}\text{C}$ (1200°F): from 649°C to 240°C
(1200°F to 464°F)

The thermal fluid is limited in temperature to a maximum:

$t_{out} = 316^{\circ}\text{C}$ (600°F)

Its specific heat at that temperature is $3.01 \text{ kJ/kg}, ^{\circ}\text{C}$ ($0.72 \text{ Btu/lb}, ^{\circ}\text{F}$). A typical volumetric flow of 5.05 l/s (80 gpm), with a density of 0.69 kg/l (5.76 lb/glm), corresponds to a mass flow rate through the dryer of 3.48 kg/s (7.68 lb/s). The inlet temperature of the thermal fluid will be 258°C (497°F) for an energy input of 2.16 GJ/h (2.05 M Btu/h). The temperature increase of the thermal fluid is 57.2°C (103°F).

Now, for a gas inlet temperature of 982°C (1800°F), the mean logarithmic temperature difference, ΔT_m , is $\Delta T_m = 220.2^{\circ}\text{C}$ (428.5°F). If the gas inlet temperature of 649°C (1200°F), the gas outlet temperature cannot be lowered to 240°C (464°F), since the thermal fluid temperature should be maintained above 260°C (500°F).

The heat exchanger thermal flux $\dot{E}$, in GJ/h (Btu/h), is of the form $KA\Delta T_m$, with K the heat transfer coefficient, in $\text{GJ/h}, \text{m}^2, ^{\circ}\text{C}$ ($\text{Btu/h}, \text{sq ft}, ^{\circ}\text{F}$) and A the heat transfer surface area, in m^2 (sq ft). To maintain the same ΔT_m as for $t_{in} = 982^{\circ}\text{C}$ (1800°F), when $t_{in} = 649^{\circ}\text{C}$ (1200°F), would require a gas outlet

temperature at 324°C (615°F). According to a manufacturer's quote, 329°C (625°F) could be achieved in practice. However, the efficiency would then be lowered to 49.2%, resulting in a 32% increase of the charge rate of the combustor, and possibly the need to choose the "next size up" for the modular incinerator.

One of the quotes was for a vertical gas flow exchanger, with a horizontal arrangement of the thermal fluid tubes. The unit was 4.6 m (15 ft) long, 1.4 m (4.5 ft) wide, 1.5 m (5 ft) high. Inlet and outlet gas temperatures were 649°C (1200°F) and 329°C (625°F), respectively.

The other design had a similar geometry. However, in this case the materials of construction were chosen to withstand the 982°C (1800°F) incoming temperature of the combustion gases. A number of rows of tube on the downstream side of the gas flow were equipped with fins to enhance heat transfer. Outlet temperature was 343°C (650°F), resulting in a heat recovery efficiency of 66.4%. The cross-sectional area was 0.6 m x 0.9 m (2 ft x 3 ft), and the length 2.15 m (7 ft).

A check was made, using the characteristics of the combustion gases at 982°C (1800°F), obtained in Appendix A1 and Chapter 3.

Outlet temperatures of 982°C (1800°F) require 7.05 kg air/kg of fuel or 7.92 kg gaseous products/kg of fuel. In a combustor of 80% efficiency, combined with a heat exchanger of 66.4% efficiency (overall efficiency: 53.1%), the amount of fuel required for the net transfer to the thermal fluid of 2.162 GJ/h (2.05

M Btu/h) is

$$\frac{2.162 \times 10^9}{.8 \times .664 \times 13.52 \times 10^7} = 301 \text{ kg/h (664 lb/h) of fuel.}$$

This requires 2125 kg/h (4681 lb/h) of air, and 2384 kg/h (5259 lb/h) of gaseous products are exhausted. Assuming that the spacing between tubes in a horizontal row is 1.3 times the outside diameter, using the terminology of reference (21):

- available section for gas flow: 0.565 of area = 0.315 m² (3.39 sq ft)
- mass velocity of gas: 7568 kg/m², h (1551 lb/sq ft, h).

The heat transfer coefficient K is the product of

- the basic convection conductance coefficient:
1.592 MJ/m², h, °F (78 Btu/sq ft, h, °F) for the
above mass velocity
- the physical properties factor: 0.135 (8.4% moisture
by wt, 1200°F)
- the arrangement factor: 0.97 (axial spacing of tubes =
1.5 diameter)
- the depth factor: 1 (depth of more than 10 rows of
tube)

Thus

$$K = 0.208 \text{ MJ/m}^2, \text{ h, } ^\circ\text{F (10.21 Btu/sq ft, h, } ^\circ\text{F)}$$

The tube area required is 43.58 m² (468.5 sq ft). If a 10% reduction in area is effected by the use of disc fins, the net tube area required is 39.26 m² (422 sq ft). In the 0.9 m x 0.6 m (3 ft x 2 ft) horizontal section, there will be 15 tubes in each

row, and 36 rows, for a total tube surface of 39.44 m^2 (424 sq ft). The height of the exchanger is approximately 2.3 m (7.5 ft), in good agreement with the manufacturer's quoted value.

COSTING OF THE UNIT

Cost estimates given by manufacturers were in a relatively narrow range of \$51,000 to \$57,000. A waste heat boiler would cost approximately \$40,000, although this figure would be increased by the cost of feed water conditioning and other steam system auxiliaries.

In assessing the economical viability of RDF use in drying glass, a figure of \$54,000 for the cost of the heat exchanger was adopted.

CHAPTER 7

ECONOMIC CONSIDERATIONS

GENERAL

In this chapter, an evaluation is made to assess as realistically as possible the economic feasibility of drying the froth-floated recovered glass using RDF as the fuel substitute, in lieu of oil.

To do this, fuel savings by not using oil have to be balanced against capital, operating and maintenance cost of the RDF combustor-heat exchanger-dryer package. The possible negative value of the substitute fuel (RDF), i.e., the avoidance cost of landfilling, is another factor taken into consideration.

The following elements will be considered:

- (a) Cost of RDF combustor-heat exchanger-dryer package: capital, installation, operation and maintenance, power and fuel
- (b) Site preparation (concrete foundation; pilings)
- (c) Cost of auxiliary fuel, C_{AF} , in \$/l (\$/gallon)
- (d) (Negative) cost of waste fuel, C_{WF} , in \$/Mg (\$/ton)

ASSUMPTIONS

The calculation is made assuming the following fixed set of parameters:

Feed: 4 Mg/h of wet feed @ 12% moisture.
 10 h operating day; 12 h working
 day (90% availability, 1 needed for
 startup and shutdown)

Capital costs: 10 year, 8% interest rate. $k_c =$
 capital recovery rate = \$0.149

Installation costs: Combustor/waste heat boiler or heat
 (includes freight) exchanger: manufacturers' data for k_{CH}
 (fraction of equipment cost)
 Dryer (thermal fluid): k_D (fraction of
 equipment cost) = 0.4
 Oil heater (thermal fluid): k_{DH} (fraction
 of equipment cost) = 0.2

Maintenance costs: Combustor/heat exchanger: $k_{MCH} = \$0.72/\text{Mg}$
 (\$0.65/t) of fuel waste (manufacturers'
 data)
 Dryer: $k_{MD} = 0.05$ (fraction of equipment
 cost for maintenance, per year)
 Oil heater: $k_{MQ} = 0.05$

Operating costs: Combustor/heat exchanger:
 - Labor: 0.5 man (manufacturer's estimate)
 $k_{LCH} = 0.5 \text{ man-h/working day}$
 - Fuel: 2.116 J (2 M Btu0 / working day
 (1 h total: startup and shutdown)
 $k_{DAF} = 2.11 \text{ GJ (2 M Btu) / day}$
 - Electricity: $k_{LCH} = 5 \text{ kWh/working h}$
 @ 3¢/kWh

Dryer:

- Labor: 0.1 man

$$k_{LD} = 0.1 \text{ man-h/working h}$$

- Electricity: $k_{ED} = 4 \text{ kWh/working h}$

@ 3¢/kWh

Oil heater:

- Labor: 0.1 man

$$k_{LOH} = 0.1 \text{ man-h/working h}$$

- Electricity: $k_{EOH} = 1 \text{ kWh/working h}$

@ 3¢/kWh

Site Preparation:

This includes the cost of a concrete slab, 30 cm (12 inch) thick and four piles.

Combustor/heat exchanger/dryer assembly

Load: 45,400 kg (100,000 lb)

Dimensions: 10.7 m x 3.7 m x 0.3 m

(35 ft x 12 ft x 1 ft)

Estimated cost: $C_{SP}^{(1)} = \$10,000$

Oil heater/dryer assembly

Load: 15,400 kg (34,000 lb)

Dimensions: 7.6 m x 3.7 m x 0.3 m

(25 ft x 12 ft x 1 ft)

Cost: $C_{SP}^{(2)} = \$8,200$

Cost of auxiliary fuel

The auxiliary fuel is assumed to be #2 fuel oil having the following properties:

Higher heating value: 45.49 MJ/kg

(19,975 Btu/lb; 143,000 Btu/gln)

Specific mass: 876 kg/m³ (7.305 lb/gln)

Cost: 10.57¢/ℓ (40¢/gln) or

\$2.65/GJ (\$2.80/M Btu) or

a parameter

Cost of waste fuel: The waste fuel is assumed to have a negative cost, C_{WF} (\$/Mg), representing the incremental cost of disposal at the recovery site, after adjustment for the ash content (as-received). Thus

$$C_{WF} = 2.83 (1 - 0.07) = 2.63^* \text{ ($/Mg of waste fuel) or a parameter}$$

COMPARISON: WASTE FUEL VS. OIL

The table below compares, on the basis of the above data,

- Case I: assembly consisting of a modular starved air incinerator, burning approximately 318 kg/h (700 lb/h of waste fuel; a heat exchanger (combustion gases to thermal fluid); a thermal fluid dryer. Net Btu input into the thermal fluid is 2.16 GJ/h (2.05 M Btu/h)
- Case II: assembly consisting of an oil heater unit delivery 2.16 GJ/h (2.05 M Btu/h) into a dryer, and a thermal fluid dryer.

Calculations are detailed for clarity.

*Values for New Orleans (Recovery 1) and fuel "B".

TABLE 7-A. COMPARISON OF COSTS FOR WASTE FUEL USE VS. OIL USE

<u>CASE I</u>		<u>CASE II</u>	
Modular Incinerator/Heat Exchanger/Thermal Fluid Dryer, Using <u>Waste Fuel</u>		Oil Heater/Thermal Fluid Dryer, Using <u>#2 Fuel Oil</u>	
<u>Equipment Costs:</u>		<u>Equipment Costs:</u>	
Modular Incinerator	\$ 90,000 (1)	Oil Heater	\$ 32,000 (1)
Heat Exchanger	54,000 (2)		- (2)
Dryer	80,000 (3)	Dryer	80,000 (3)
<u>Installation Costs:</u>		<u>Installation Costs:</u>	
Incinerator + Heat Exchanger ($k_{CH} = .28$)	40,000 (4)	Oil Heater	6,400 (4)
Dryer ($k_D = 0.4$)	32,000 (5)	Dryer	32,000 (5)
<u>Site Preparation:</u>		<u>Site Preparation:</u>	
Foundation & Piles, C_{Sp}	<u>10,000 (6)</u>	Foundation & Piles, C_{Sp}	<u>8,200 (6)</u>
Total Front-End Costs, (1) thru (6)	<u>\$306,000 (7)</u>	Total Front-End Costs, (1) thru (6)	<u>\$158,600 (7)</u>
<u>Yearly Costs</u>		<u>Yearly Costs</u>	
Note:		See note in column I	
Glass Production = 4 Mg/h x 10 (operating hours) = 40 Mg/day = 10^4 Mg/year (1 yr = 250 D)			
1 working day	10 operating hours 12 working hours		

TABLE 7-A (Continued)

<u>CASE I</u>	
Front-End Cost, Princ. & Interest (.149 x (7))	\$ 45,600 (8)
<u>O & M</u>	
. Labor: $k = k_{LCH} + k_{LD} = 0.6$ (man)	
0.6 man x $\frac{12 \text{ working hours}}{8 \text{ hr (shift)}} = 0.9$ (man)	
0.9 x \$20,400	18,360 (9)
. Fuel Oil (Auxiliary)	
2 M Btu x 2.80 x 250 (days)	1,400 (10)
. Fuel (Waste)	
3.18 x 2.63 x 250	(2,090) (11)
. Electricity: 5 kWh (combustor/ H.E.), 4 kWh (dryer)	675 (12)
. Maintenance:	
- Combustor/Heat Exchanger ($k_{LCH} = 0.65$ \$/Mg of waste), preventive maintenance 0.65 x 3.5 x 250	570 (13)
- Dryer, $k_{MD} = 0.05$ 0.05 x 80,000	4,000 (14)
Total Yearly Cost (8) thru (14)	\$68,515 (15)

<u>CASE II</u>	
Front-End Cost, Princ. & Interest (.149 x (7))	\$ 23,630 (8)
. Labor: $k = k_{LOH} + k_{LD} = 0.2$ (man)	
0.2 x $\frac{12}{8} = 0.3$ (man)	
0.3 x \$20,400	6,120 (9)
. Fuel Oil (Oil Heater)	
$\frac{2 \text{ M Btu}}{.85 \text{ (efficiency)}} \times 2.80 \times 10 \text{ (hr)} \times 250$	16,470 (10)
-	(11)
. Electricity: 4 kWh (dryer) 4 x .03 x 10 x 250	300 (12)
. Maintenance:	
- Oil Heater, $k_{MOH} = 0.05\%$ 0.05 x 32,000	1,600 (13)
- Dryer, $k_{MD} = 0.05$ 0.05 x 80,000	4,000 (14)
Total Yearly Cost (8) thru (14)	\$52,120 (15)

TABLE 7.A (continued)

<u>CASE I</u>		<u>CASE II</u>	
<u>Auxiliary Calculations</u>		<u>Auxiliary Calculations</u>	
Cost per Mg of Glass Processed	\$6.85	Cost per (Short Ton) of Glass Processed	\$5.21
<u>Breakeven Case</u>		Fuel Cost per (Short) Ton Processed	\$1.64
Price of Oil: 22.08¢/liter (83.69/gln)			
Total Yearly Cost (I or II)	\$70,040		
Cost per (Short) Ton of Glass Processed	\$7.00		

Comments

It is seen from Table 7-A, that at present prices of oil, it does not pay off to use RDF as the substitute fuel for oil.

The reasons are apparent:

- the total installed cost of the combustor/heat exchanger combination (\$185,800) is approximately five times higher than the installed cost \$38,400) of an oil heater delivering the same amount of energy to the dryer.
- the quantity of waste fuel needed for the drying process is so small (about 0.35 Mg/h) that the impact of the negative cost of the fuel (land-fill avoidance), item (11) of Table 7-A, is negligible. It represents only 3.1% of the yearly costs, and 12.7% of the amount (\$16,395) by which it is cheaper to operate with assembly II, at present fuel oil prices.
- at present prices of the equipment, labor, electricity and incremental disposal, the breakeven situation would occur, between cases I and II, if the present price of oil is multiplied by a factor, F_{oil} , such that the incremental oil savings in items (10) (cases I and II),
$$(F_{oil} - 1) (16,470 - 1,400) = 15,079 (F_{oil} - 1)$$
balance the yearly costs (item (15), cases I and II):
$$68,515 - 52,120 = 16,395$$

Thus

$$F_{oil} = 2.09 \text{ } (\$6.17/\text{GJ or } \$5.85/\text{M Btu})$$

If this value is taken as an average at the equipment mid-life (five years), breakeven would be reached only if oil prices more than quadruple over the next ten years.

CONCLUSIONS

In conclusion, drying recovered froth-floated glass by using RDF at the recovery site does not appear to be an economically viable alternative to the use of fuel oil to heat a dryer's thermal fluid. It would be a marginally interesting proposition only if oil prices were to increase more than four-fold over the next ten years.

Further investigation of the proposed scheme would have to be recommended only as a technological experiment in use of on-site fuels at recovery sites, not on its economic merits. The elements of a testing program are outlined in the following chapter.

CHAPTER 8

OUTLINE OF TEST PLAN

GENERAL

A test program is outlined to provide needed experimental data on the drying of froth-floated glass using a solid waste fuel.

It is assumed that the assembly to be tested comprises:

- (1) a modular incinerator-heat exchanger assembly, using a waste fuel (such as fuel "B", Chapter 3) to heat a thermal fluid to 316°C (600°F) (max.).
- (2) an indirect, thermal fluid dryer of the hollow screw type.

At the site used in this discussion (Recovery 1, New Orleans), such a dryer is already installed. The energy required for the glass drying is provided at present by an oil-heater, fired with fuel #2, and capable of delivering 1.32 GJ/h (1.25 M Btu/h).

If, at the noted point, the efficiency of the hollow screw dryer is taken to be 75%, interpolation from Table 4-B gives

- (a) - for the energy input required per unit mass of wet feed at 12% initial moisture:
507 kJ/kg (218 Btu/lb)
- rated throughput at 1.32 GJ/h (1.25 M Btu/h) from the heater: 2.6 Mg/h (5,734

lb/h)

- (b) - for the energy input required per unit
mass of wet feed at 75% initial moisture:
356 kJ/kg (153 Btu/lb)
- rated throughput at 1.32 GJ/h (1.25 M
Btu/h) from the heater: 3.71 Mg/h (8,169
lb/h)

OBJECTIVES OF THE TESTS

The test would determine technical and operating data when drying recovered froth-floated glass using a solid waste fuel recovered at the site. More specifically, it would investigate

- the combustion efficiency, η_c
- the heat exchanger efficiency, η_{he}
- the dryer efficiency, η_d

at various levels of initial and final moisture of the product, for various throughputs and waste fuels.

The data to be generated are given below.

DATA TO BE GENERATED

The reader should refer to Fig. 8-1, where existing hardware and proposed additions are represented, and Fig. 8-2 for test plan data point reference. In Fig. 8-2, the data sampling points are represented by small lettered circles ($\dot{m}$ - mass flow rate, ω - moisture content, T - temperature, P - pressure). Arrowheads indicate directions, and dashed lines indicate control feedback loops.

Fig. 8-1. Existing Recovery 1 Glass Drying Module with Proposed Modifications

Fig. 8-2. Drying Circuit

1. Mass Flow Rate of Wet Feed (Fig. 8-2)

The glass slurry is conveyed from the froth flotation tank to the hydrocyclone via a pump. It is gravity fed from the hydrocyclone to the vacuum filter. From the vacuum filter, it is scroll fed to the dryer. A chute is provided at the exit of the dryer to collect and weigh the dry fuel on a periodic basis. Maximum de-watering of the glass occurs just prior to entering the scroll feeder. If the moisture content of the product is sampled just prior to entering the scroll feeder, and the flow rate and moisture content of the product are sampled at the dryer exit, the mass flow rate of wet feed entering the dryer can be determined.

2. Mass Flow Rate of Thermal Fluid

Given that the density of the thermal fluid is known at the operating temperature of the system, and the geometry of the system is fixed, the mass flow rate of the thermal fluid can be determined with a turbine flow meter. A frequency converter is required to measure output from the flow meter. Conversion of frequency data to flow rate data can be done manually using simple precalibrated charts. Since the flow rate will not be changing rapidly with time, periodic sampling will be sufficient. The maximum flow available for the dryer is 5.05 l/s (80 gpm).

3. Temperature of Thermal Fluid into Dryer (Fig. 8-2)

This parameter will have to be monitored continuously, say with a Chromel-alumel thermocouple. The signal will be sent simultaneously to an indicator which may be recorded manually on a periodic basis, and to a controller which can continuously

regulate the flow of flue gases through the heat exchanger by directing any or all of the flue gases directly to the stack.

4. Temperature of Thermal Fluid out of Dryer (Fig. 8-2)

A thermocouple connected to an indicator, monitored manually on a periodic basis, will record this parameter. If a selector switch is provided, the input and output thermocouple can be monitored with the same indicating device.

5. A feedback loop will have to be provided from the output of the thermal fluid flow meter to the heat control circuit. The purpose of the signal will be to direct the flue gases to the stack when the flow of the thermal fluid falls below a safe minimum value.

A pressure sensitive switch downstream of the thermal fluid pump will perform the same task probably more economically.

6. Mass Flow Rate of Air into Combustor

Wet/dry bulb temperature and manometer are sampled manually.

7. Mass Flow Rate of Fuel into Combustor

It is expected that the fuel will be batch-fed to the metering bin. The moisture content and overall weight can be sampled manually prior to loading the metering bin.

8. Mass Flow Rate of Ash Leaving Combustor

The ash will be batch-hauled to landfill and may be sampled prior to landfilling.

If the retrofit is restricted to off-the-shelf items of proven technology, the heat exchanger chosen for this application will be restricted to a heat recovery system (flue gas to thermal fluid).

ESTIMATED COST OF INSTRUMENTATION

The cost of the required instrumentation is listed below in Table 8-A on an itemized basis.

PROPOSED LAYOUT

A proposed layout, showing the existing and proposed equipment, is shown in Fig. 8-3. In this layout, the heat exchanger is located on a part of an existing concrete pad directly adjacent to the modular incinerator stack. The inlet and outlet flanges of the thermal fluid circuit in the heat exchanger are also conveniently located near the existing indirect dryer.

The oil heater system would remain at its present location and be used as a standby.

The metering bin used for the waste fuel would be located on the roadside for ease of delivery of the fuel fractions recovered on the site.

TABLE 8-A. REQUIRED INSTRUMENTATION

Qty.	Item	Function and Type	Estimated Price
1	Flow meter	Thermal fluid mass flow rate	\$1,500.00
1	Wet/dry bulb	Humidity ratio, air leaving the dryer	700.00
2	Thermocouples	Thermal fluid temperature (in/out)	
2	Thermocouples	Glass temperature (in/out)	1,520.00 (complete with probes & wires)
1	Flat Bed Recorder/Indicator	Sample or monitor all functions	
1	Selection switch	Eliminate need for more than one recording device	100.00
1	Signal conditioning unit	Normalize all analog signals to recorder range	400.00

Minimum slab size: 12' wide x 35' long
 Pilings required: 4 (best estimate before survey)

Fig. 8-3. Glass Plant Layout with RDF Combustor

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APPENDIX A1

RDF: COMBUSTION AIR,

PRODUCT GAS COMPOSITION, TEMPERATURES

COMBUSTION CALCULATIONS (FIRST EXAMPLE FUEL)

The refuse-derived fuel considered in the calculation is a representative example described in the body of the text (Table 2-A).

(a) Stoichiometric combustion

The method followed for the determination of the stoichiometric combustion air is as described by Corey (21). Standard air is defined as air containing 1.3%, by dry weight, of moisture. Dry air is assumed to have the composition, by volume: O = 21%, N₂ = 79% (including inerts).

To compute the two adiabatic combustion temperatures, the higher heating value was corrected for the moisture produced by the combustion of the fuel. From the composition of combustion products (gaseous products and ash), the specific heats at atmospheric pressure, $\bar{c}_p$ (all products) and $\bar{c}_{p,g}$ (gaseous products) were computed iteratively at mid-range over the range of temperatures between ambient and combustion temperatures, from the values of c_p for the various components (22).

Results are summarized in Tables A-I and A-II.

(b) Non-stoichiometric combustion

Similar calculations were performed for various values of the excess air, EA, and are summarized in Tables A-I and A-II.

TABLE A-I. COMBUSTION PARAMETERS VS. EXCESS AIR

EXCESS AIR, E.A.

		EA = 0.	EA = 0.25	EA = 0.50	EA = 0.75	EA = 1.0
kJ/kg, °C (kcal/kg °C or Btu/lb °F)	$\bar{c}_{p,g}$	1.361 (0.325)	1.285 (0.307)	1.243 (0.297)	1.197 (0.286)	1.172 (0.280)
	$\bar{c}_p$	1.348 (0.322)	1.277 (0.305)	1.235 (0.295)	1.193 (0.285)	1.168 (0.279)
Equivalent mol. wt.		28.66	28.68	28.69	28.70	28.71
T* _{EA} (adiabatic combustion temp) °C (°F)		1547 (2817)	1357 (2475)	1187 (2169)	1086 (1986)	985 (1805)
kg air/kg fuel		4.83	6.04	7.32	8.45	9.66
T _g , °C (°F)		1242 (2267)	1090 (1993)	954 (1748)	873 (1603)	792 (1458)

TABLE A-II. COMPOSITION OF GASEOUS PRODUCTS VS. EXCESS AIR

Component	EXCESS AIR, E.A.									
	EA = 0.		EA = 0.25		EA = 0.50		EA = 0.75		EA = 1.0	
	Vol. Percent	Wt. Percent	Vol. Percent	Wt. Percent	Vol. Percent	Wt. Percent	Vol. Percent	Wt. Percent	Vol. Percent	Wt. Percent
CO ₂	15.11	23.19	12.47	19.16	10.62	16.29	9.25	14.18	8.19	12.55
H ₂ O	19.28	12.11	16.29	10.24	14.19	8.91	12.61	7.91	11.41	7.15
N ₂	65.51	64.45	67.58	66.44	69.02	67.84	70.13	68.90	70.96	64.70
O ₂	0.00	0.00	3.57	3.95	6.09	6.79	7.95	8.87	9.39	10.47
SO ₂	0.10	0.25	0.09	0.21	0.08	0.17	0.06	0.14	0.05	0.13

(c) Non-adiabatic combustion

A combustion gas temperature, T_g , was computed according to equation (3-1) and is given for each case on the last line of Table A-I.

COMBUSTION PARAMETERS FOR EA = 0.46 ($T_g = 982^\circ\text{C}$ or 1800°F)

The composition of flue gases and other relevant quantities are given below for the value EA = 0.46 corresponding to a gas temperature $T_g = 984^\circ\text{C}$ (1803°F).

<u>Component</u>	<u>Volume Fraction, Percent</u>	<u>Weight Fraction, Percent</u>
CO ₂	10.89	16.71
H ₂ O	14.49	9.10
N ₂	68.91	67.74
O ₂	5.63	6.27
SO ₂	0.08	0.18

$\bar{c}_p$ (kJ/kg, °C)	1.2367
$\bar{c}_{p,g}$ (kJ/kg, °C)	1.243
Equiv. Mol. Wt.	28.7
T_{EA}^* , °C	1224.
kg air/kg fuel	7.05
T_g , °C	984.

COMPOSITION AND DATA FOR DILUTED COMBUSTION GASES AT 649°C

(1200°F)

<u>Component</u>	<u>Volume Fraction, Percent</u>	<u>Weight Fraction, Percent</u>
CO ₂	6.69	10.26
H ₂ O	9.71	6.09
N ₂	72.18	70.87
O ₂	11.37	12.67
SO ₂	0.05	0.11

Equiv. Mol. Wt. 28.72

 $\bar{c}_{p,g}$ (kJ/kg °C) 1.185

FUELS B AND C': BREAKDOWN BY COMPONENTS

Fuel B

<u>Component</u>	<u>Wt, percent</u>	<u>Moisture,* percent</u>	(1) <u>Dry fraction before magnetic recovery, percent</u>
Paper	58.4	20	46.7
Yard wastes	4.9	50	2.5
Plastics	4.9	10	4.4
Other heavy organics	25.4	50	12.7
Ferrous	4.9	-	4.9
Aluminum	0.6	-	0.6
Other non- ferrous	-	-	-
Glass	0.3	-	0.3
Stone	-	-	-

Breakdown by Components (continued)

<u>Component</u>	<u>Wt, percent</u>	<u>Moisture*, percent</u>	(1) <u>Dry fraction before magnetic recovery, percent</u>
Fines	<u>0.6</u> 100.0	-	<u>0.6</u> 72.0

<u>Component</u>	(2) <u>Dry fraction after magnetic recovery, percent</u>	(3) <u>Percentage, by wt of (2)</u>	(4) <u>HHV of dry material, Btu/lb</u>
Paper	46.7	68.7	(7,500)
Yard wastes	2.5	3.7	(8,000)
Plastics	4.4	6.4	(15,000)
Other heavy organics	12.7	18.7	(9,000)
Ferrous	0.2	0.3	
Aluminum	0.6	0.9	
Other non-ferrous	-	-	
Glass	0.3	0.4	
Stone	-	-	
Fines	<u>0.6</u>	<u>0.9</u>	
	68.0	100.0	

*by weight of component

Estimated ash content of components (by dry weight):

Paper	9%
Yard wastes	4%
Other heavy organics	5%

Note: Other non-ferrous, glass, stones and fines are assumed to report to the ash fraction.

Higher heating values are: $(HHV)_{B,dry} = 18.82 \text{ MJ/kg (8092 Btu/lb), dry basis}$

$(HHV)_B = 13.14 \text{ MJ/kg (5778 Btu/lb), as-received}$

Ash contents are: $(Ash)_{B,dry} = 9.2\%, \text{ dry basis}$

$(Ash)_B = 6.6\%, \text{ as-received}$

Moisture: 28.6%

Fuel C'

Based on a breakdown by component, fuel "C", obtained as described in Chapter 3, is found to have a large percentage of fines (<1/4 inch) and a higher heating value, as received:
 $HHV = 9.50 \text{ MJ/kg (4085 Btu/lb)}$.

An alternate fuel, C', can be obtained from C, after a 90% efficient screening of all "stones and fines" (under 1/4 in). In this case the composition is:

<u>Component</u>	<u>% by wt, dry</u>
Paper	54.7
Yard wastes	18.4
Glass	7.5
Plastics	3.6
Stones and fines	6.2
Heavy organics	9.3
Aluminum	-
	100.0
	Moisture = 29.5%

and the heating values are

C': $(HHV)_{dry} = 18.88 \text{ MJ/kg (8117 Btu/lb)}$

C': $(HHV)_{C'} = 11.67 \text{ MJ/kg (5017 Btu/lb)}$

Using the same procedure as above, the ash contents would be

C': ash content, dry basis = .15.4%

C': ash content, as-received basis = 10.9%

Moisture: 29.5%

APPENDIX A2

TECHNICAL DATA ON DRYING PROCESS AND DRYERS

DRYING PROCESS AND DRYING EFFICIENCY

Drying, the removal of moisture from a solid substance, is a basic operation in materials processing. The use of the term is generally confined to cases where the removal of the moisture is by evaporation into a gas stream, and thus would not apply to adsorption, mechanical dewatering, centrifugation, etc. (23).

The present discussion is further limited to air, or combustion gases as the agent used to remove the moisture. Also, the word "moisture" will also mean "water" or "water vapor," as the case may be.

The moisture content of the solid being discussed can be expressed on a wet (subscript "w") or dry (subscript "d") basis, as

$$f_w = \frac{\text{mass of moisture}}{\text{mass of wet solid}} \quad (\text{per unit})$$

$$x_d = \frac{\text{mass of moisture}}{\text{mass of dry solid}} \quad (\text{per unit})$$

If f_s is the weight fraction of solids in the wet feed product, the following hold:

$$f_w + f_s = 1$$

$$x_d = \frac{f_w}{1 - f_w}$$

$$f_w = \frac{x_d}{1 + x_d}$$

The operation can be achieved in a batch mode, or continuously. The latter mode would normally be preferred for on-line processing, at a resource recovery site, of a constant throughput of glass, in the order of several tons per hour, over a 12-hour day.

The heat is conveyed into the dryer by a heat conveying medium, which can be hot air, hot combustion gases, a thermal fluid, or steam. This heat conveying medium might be in direct contact with the substance to be dried (direct drying) or the heat might be transferred indirectly, for instance through tubes, hollow-spiral flights or other conducting surfaces, by infrared radiation or dielectric heating (indirect drying) (Fig. 1-2).

If no heat is supplied within the dryer nor lost to the surroundings, the drying process is adiabatic. The heat conveying medium then transfers some of its sensible heat to the medium to be dried, mostly as latent heat if vaporization of the moisture removed but also as sensible heat of the solids.

If the hottest conveying medium is first transferring heat to the direct solids, i.e., in a countercurrent arrangement, some advantage might be gained in effectively removing bound moisture, or that part of the moisture exerting a partial pressure lower than that of the liquid at the given temperature. Drawbacks are the high sensible heat carried out of the dryer by the dry solids, which entails an energy loss, and the risk of damaging the material being dried. The latter point especially should be of

concern when drying ordinary glass, a material which softens enough to lose its shape at about 600°C (1100°F).

In the co-current arrangement, the hottest conveying medium transfers heat to the solids containing the highest percentage of moisture. While the surface moisture vaporizes, the temperature of the solid will not exceed the wet-bulb temperature of the gas carrying away the moisture.

The thermodynamic process in the dryer occurs at essentially constant (atmospheric) pressure. Thus, if q_{AB} is the heat per unit mass given by the heat conveying medium (HCM) between the boundaries A (inlet, 0) and B (exit, 1) of the system:

$$q_{AB} = h_1 - h_0$$

In the above, h is the enthalpy per unit mass. For constant (or averaged) c_p over the temperature range (t_0 , t_1)

$$q_{AB} = c_p (t_1 - t_0)$$

The total enthalpy flux from the HCM into the dryer is noted $\dot{Q}_{AB}$, and $\dot{m}$ is the mass flow rate of the HCM. In a perfectly adiabatic dryer, all of this energy flux would be transferred:

- to the medium to be dried, in the form of latent heat of vaporization needed to release the moisture in the gaseous stream which carries it away. If $\dot{m}_\ell$ is the mass flow rate of liquids into (subscript "in") or out of (subscript "out") the dryer, this term, noted $\dot{Q}_{vap,\ell}$, can be expressed as a function of the latent heat of vaporization

ℓ_{vap} at the given pressure and a temperature of 100°C (a conservative assumption):

$$\dot{Q}_{\text{vap},\ell} = \dot{m}_{\ell} \ell_{\text{vap}}$$

$$\ell_{\text{vap}} = 2,257 \text{ kJ/kg (539.1 kcal/kg; 970.3 Btu/lb)}$$

- to the liquid heated up to the temperature at which vaporization occurs, t_{vap} . The corresponding term, called $\dot{Q}_{h,\ell}$, is

$$\dot{Q}_{h,\ell} = c_{p,\ell} (t_{\text{vap}} - t_{\ell,\text{in}}) \dot{m}_{\ell}$$

$c_{p,\ell}$ is the specific heat of the liquid at constant pressure.

- to the solids heated up from the inlet temperature $t_{s,\text{in}} = t_{\ell,\text{in}}$ to the exit temperature $t_{s,\text{out}}$. This is a term

$$\dot{Q}_{h,s} = c_{p,s} (t_{s,\text{out}} - t_{s,\text{in}}) \dot{m}_s$$

- to the gaseous stream carrying the moisture out of the dryer (if the drying is indirect)

$$\dot{Q}_{h,g} = (c_{p,g,\text{out}} t_{g,\text{out}} - c_{p,g,\text{in}} t_{g,\text{in}}) \dot{m}_g$$

The subscript g refers to the gas. In an indirect dryer, the "gas" is a draft (natural or forced) of air progressively warming up and collecting the moisture released in the process.

For the perfectly adiabatic dryer, the enthalpy balance reads

$$\dot{Q}_{\text{AB}}^* = \dot{Q}_{\text{vap},\ell} + \dot{Q}_{h,\ell} + \dot{Q}_{h,s} + \dot{Q}_{h,g}$$

Allowing for losses, at rate $\dot{Q}_L$, the previous

equation is modified to read

$$\dot{Q}_{AB} = \dot{Q}_{\text{vap},l} + \dot{Q}_{h,l} + \dot{Q}_{h,s} + \dot{Q}_{h,g} + \dot{Q}_L$$

These losses allow for radiation to the outside, and convective and conductive exchange to the surroundings, and are determined empirically from test data. The efficiency of the dryer can be defined in a variety of ways. Here it will be taken as the ratio of: the heat strictly needed to vaporize the moisture, heat up the liquid to the vaporization temperature, the solids and gas removing the moisture to their exit temperature, to the sensible heat input (referred to ambient temperature, 20°C) into the dryer from the heat conveying medium, noted H_{in} . Thus, for the direct dryer

$$\eta_d \equiv \frac{\dot{Q}_{AB} = \dot{Q}_L}{H_{in}} \equiv \frac{q_{amb}^*}{c_p (t_1 - t_{amb})}$$

$$q_{amb}^* \equiv \frac{\dot{Q}_{AB} \text{ (for } \dot{Q}_{h,g} \equiv 0)}{m}$$

In the case of indirect drying, the enthalpy content of the heat conveying medium is not lost to the surroundings. Thus at the dryer exit

$$H_{in} = c_p (t_1 - t_0) \dot{m}$$

and

$$\eta_d = \frac{\dot{Q}_{AB} - \dot{Q}_L}{c_p (t_1 - t_0)\dot{m}} = \frac{\dot{Q}_L}{c_p (t_1 - t_0)\dot{m}}$$

The above expressions for η_d are used as the dryer efficiency for direct and indirect drying, respectively.

GEOMETRY AND TECHNICAL FEATURES OF MAIN TYPES OF DRYERS

Direct Rotary Dryers

The direct rotary dryer has the geometry represented in Figure A2-1. It has the shape of an elongated cylinder, rotating about its axis of symmetry, and traversed (co-current or counter-current) by a stream of hot air or combustion gases (24). The solids, fed at one of the ends, progress toward the other end at a slow rate, the axis being very slightly inclined to the horizontal. Fill ratios, or percentage of the section occupied by the material to be dried (at rest), seldom exceed 20% (25).

Heat transfer to individual particles, and vaporization of the free moisture is enhanced by the rotation, which keeps the material tumbling. In many instances, it is also helped by lifters which bring the particles to a high point in the section (3/4 to 1 diameter height). Depending on the angle of the lifters and the material's angle of repose, at some location near the top of the section, the particles are dropped as a solids shower and are well exposed to the drying gas.

A direct dryer design, broadly speaking, will be characterized by a set of parameters:

- Diameter D: This dimension directly affects the height from which the material carried by the lifters showers down. Rotary dryers exist in a range of diameters from as little as 0.15 m (6 inch) to as much as 7.32 m (24 ft). However, practical designs of shop fabricated units to be transported and assembled on the site will limit D to approximately 3.66 m (12 ft). A throughput requiring a larger diameter is split between two parallel units (26).
- L/D: length to diameter ratio. For most practical designs, L/D appear to be in the 3 to 7 range.
- Ω , rate of rotation: Peripheral speeds are in the range 0.20 to 0.51 m/s (40 to 100 FPM).
- Lifters: extend 8 to 12% of the diameter. Their number is in the range 2D to 3D (D in feet) (27).
- Heat transfer rate per unit volume of dryer: since the "heat transfer surface," i.e., the external surface of exposed solids, is not exactly known or measurable, it is preferable to use the heat transfer rate per unit volume of the dryer as a measure of compactness of the device for a given logarithmic mean temperature difference between the gases and the solids. Since this temperature varies over a wide range, 167°C to 1670°C (300 to 3000°F) in individual applications, this parameter is comprised in a wide range from 37.2 to 223.2 MJ/m³,h (10³ to 6 x 10³ Btu/cft, h).

Fig. A2-1. Geometry of Direct Dryer

Indirect Dryers

As previously discussed, the indirect dryer uses a heat conveying medium (HCM) not directly in contact with the solids to be dried. The moisture liberated by the material being dried is carried out of the dryer enclosure by means of a natural or forced draft of air, which may or may not be preheated to increase its moisture carrying capacity.

The heat conveying medium should have a high enthalpy per unit volume transversing the dryer at velocities which do not create excessively high pressure drops. The enthalpy transfer rate is proportional to $\dot{m}_{HCM}$, mass flow rate of the heat conveying medium, and either ℓ_{VAP} (if there is condensation) or $c_{p,HCM} (t_{HCM} - t_s)$.

"Compact" heat transfer can thus be achieved by using

- a condensing gas (ℓ_{VAP} large), such as steam
- a hot thermal fluid or molten salt ($\rho_{liquid} \gg \rho_{gas}$ large, t_{HCM} large, c_p large)

The heat exchanger will also be more compact (per unit area of outer shell, say) if the solid surface of exchange at the interface between the HCM and the material to be dried is as large as possible.

Steam-heated Indirect Dryers

Figure A2-2 shows a cross-sectional and longitudinal schematic view of a dryer of that type. Steam tubes disposed inside the rotating shell are fed by a header into which steam is introduced through a rotary joint. Air induced by natural draft is used only to carry away the product moisture, and dusting or

Fig. A2-2. Indirect Dryer (Steam-Heated)

carryover should be insignificant. Drying efficiencies are in the 50% to 60% range.

Another manufacturer offers a very different design, in which the feed material and moisture-carrying air circulate inside symmetrically arranged tubes equipped with removal suspension blades, acting both as lifters and feeders (in the feed and air inlet chamber). Steam heats the tubes from the outside. The objective is to uniformly disperse the feed and maximize the heat transfer surface. Drying efficiency is claimed to be about 85%.

Thermal Fluid or Molten Salt Indirect Dryers (28)

In this type of equipment, the heat conveying medium is either a molten salt or a thermal fluid at temperatures in the 315°C to 370°C (600°F to 700°F).

In one configuration, the molten salt heats the dryer metal jacket, made of hollow flights of octagonal sections, from the outside. The lower fifth or so of its height is submerged in the molten salt. The HCM itself is heated by 2 to 3 loops, on either side of the vertical, through which the combustion gases are flowing. Obvious advantages of this design is that no high pressure rotary seals are needed, and that dusting and attrition of the solids is kept negligibly small by having in contact with the solids only the air needed to carry out moisture. For a drum of diameter 1.83 m (6 ft), 30.5 m² per m are obtained with the fin arrangement, and heat transfer coefficients in the order of 122.5 kJ/m², °C, h (6 Btu/sq ft, °F, h) are quoted. Thermal efficiencies (including a 75% efficiency of combustion to heat the salt) are about 55% overall, for a drying efficiency of 73%,

approximately.

Another design (Fig. A2-3) includes helicoidal hollow flights through which the HCM, a thermal fluid of high specific heat (in the range 2.51 to 2.93 kJ/kg, °C (0.6 to 0.7 Btu/lb, °F) at the operating temperature) flows. Heat transfer is from the inside (flights, hollow shaft) to the outside (materials to be dried) for the first pass. The screw conveyor, or arrangement of multiple, intermeshing screws, conveys the solids to be dried towards the exit section. Additional heat transfer area is provided by thermal fluid circulating through jackets or jacket covers in the trough containing the flights. The HCM moves down and back the length of the jacketed trough. This type of dryer exists in single, double or quadruple screw arrangements. Body length (trough) ranges from 3 m to 9.3 m (10 ft to 24 ft), and screw diameters from 18 cm to 61 cm (7 inches to 24 inches). Heat transfer coefficient (from tests on ground glass) is estimated to be about 306 kJ/m², °C, h (15 Btu/sq ft, °F, h). Dryer efficiency is expected to be in the 70% to 75% range.

Fluidized Bed Dryers

In a standard type of fluidized bed dryer, as sketched in Fig. 1-2, hot combustion gases or air are introduced in a lower plenum, then flow through orifice plates or screens to fluidize a bed of material to be dried. Such dryers can be used in the batch or continuous modes. The high surface area of exchange between the gas and the particles significantly increases heat transfer, and the conditions are fairly uniform throughout the bed. An extensive treatment of fluidized bed drying is given by

Fig. A2-3. Indirect Dryer (Thermal Fluid)

Vanecek (9). One manufacturer quoted a thermal efficiency of 60% for this type of dryer.

A configuration used by a manufacturer for drying plastics, pigments, etc., includes a baffled maze in the fluidized volume, and the product progressively dries as it advances from side to side. A heat transfer medium (hot water or steam) circulates in coils to provide conductive heat. The depth of the fluidized bed, using ambient air, is 0.6 to 0.9 m (2 to 6 ft). Heat transfer coefficients of 408 to 3062 KJ/m², °C, h (20 to 150 Btu/sq ft, °F, h) are said to be achievable. Capacities range from 91 to 22,680 kg/h (200 to 50,000 lb/h). Floor areas range from 0.3 m x 1.2 m to 3.1 m x 12.2 m (1 ft x 4 ft to 10 ft x 40 ft).

These types of dryers have to include a system for the recovery and possible reinjection of elutriated fines. This is obviously a significant drawback, compared to indirect dryers in which entrainment can be made negligible by minimizing velocities in the draft carrying away the moisture. Another serious disadvantage of direct drying by combustion gases is the contamination of glass by air particles of froth-floated glass, a product in which ceramic particles have to be kept under a few ppm by weight to meet specifications (29).

Vanecek (30) gives values for V_{mf} , the minimum fluidization velocity for sand, and V_{mf} , the velocity at which a particle will be elutriated from the bed, as a function of the particle diameter, D_p . It is seen that in the case of a polydisperse mixture of particles, ranging in diameters between D_{min} and D_{max}

$$D_{min} \leq D_p \leq D_{max}$$

velocity needed to fluidize the larger particles might cause elutriation of the smaller ones. As seen in Chapter 2, the range of sizes for recovered froth-floated glass is such that:

$$D_{\min} = 0.107 \text{ mm}$$

$$D_{\max} = 1.0 \text{ mm}$$

with a particle size distribution, by weight, given in Fig. 2-1.

From the equation given by Kunii and Levenspiel (31), and for a bed at 121°C (250°F), estimates for the minimum velocity of fluidization are found to be, in the present case:

$$V_{mf} = 0.0077 \text{ m/s for } D_p = 0.107 \text{ mm}$$

$$V_{mf} = 0.6718 \text{ m/s for } D_p = 1.0 \text{ mm}$$

A fluidization velocity of 0.67 m/s is required to fluidize the larger particles. By trial and error, a settling velocity, U_s , of 0.67 m/s corresponds to a diameter of the particle.

$$D_p^* = 0.34 \text{ mm}$$

This diameter is determined by solving for V_p the equations (for $\rho_p \gg \rho_g$)

$$1/2$$

$$U_s = \left[\frac{\rho_p}{\rho_g} \frac{4g}{3} D_p \frac{1}{C_D} \right]^{1/2}$$

$$C_D = \frac{24}{Re} \left[1 + \frac{3}{16} Re \right]^{1/2}$$

$$Re_D = \frac{U_s D_p}{\mu_g / \rho_g}$$

Since particles having settling velocities smaller or equal to the settling velocity are likely to be elutriated (32), it can be stated, by reference to Fig. 2-1, that approximately 37% (the

population of the first four classes) of the material will be carried away from the bed.

Through-circulation Dryers

Some hybrid designs combine the features of through-circulation and rotary dryers. For example, Fig. A2-4 shows a unit consisting of a revolving tapered drum fitted with louvers. The hot gases enter into the drum beneath the solids layer. Low dusting is said to result in this type of configuration (33).

Fig. A2-4. Through-circulation Dryer

APPENDIX A3
ENTRAINMENT OF FINES IN DIRECT DRYING OF
FROTH-FLOATED GLASS

This appendix evaluates the amount and size of glass particles in the froth-floated material which could be entrained in the stream of hot combustion gases used for directly drying the material.

The polydisperse mixture of glass particles has the particle size distribution shown in Fig. 2-1.

The incoming flue gases have been diluted to a temperature acceptable to the dryer construction material, i.e., approximately 649°C (1200°F). As seen in Appendix A1,

- 12.90 kg of gaseous products are produced per kg of fuel
- the wt. average molecular weight of the mixture is 28.72
- the specific heat of the combustion gases is 1.185 kJ/kg, °C or 0.283 kcal/kg, °C (Btu/lb, °F)

Case "A": inlet conditions

The thermodynamic condition of the gases at the inlet are

$$\begin{aligned}p_{\text{gas}} &= 101,320 \text{ Pa (1 atmosphere)} \\r_{\text{gas}} &= 289.4 \text{ (J/kg}^\circ\text{K)} \\T_{\text{gas}} &= 922^\circ\text{K (1200}^\circ\text{F)}\end{aligned}$$

$$\rho_{\text{gas}} = 0.38 \text{ kg/m}^2 \text{ (0.0237 lb/cft)}$$

The velocity in the initial section of the dryer is computed from the gases volumetric flow rate and a fuel input of 201.4 kg/h (444 lb/h) (HHV = 13.43 MJ/kg or 5775 Btu/lb; combustor efficiency 80%)

$$\dot{Q} = 1.90 \text{ m}^3/\text{s} \quad (4018 \text{ cfm})$$

In a typical dryer having a 1.83 m (6 ft) diameter, at 15% fill ratio, the velocity in the initial section (if uniform) is

$$V = 0.85 \text{ m/s} \text{ (2.79 ft/s)}$$

To evaluate the degree of entrainment, the free settling velocity, U_s , is computed from

$$\frac{1}{2} \rho C_D U_s^2 \pi \frac{D_p^2}{4} = \frac{\pi}{6} D_p^3 (\rho_p - \rho_g) g$$

$$C_D U_s^2 = \frac{4}{3} D_p \left(\frac{\rho_p}{\rho_g} - 1 \right) g \quad (\text{A3-1})$$

C_D , the drag coefficient, is assumed to depend on the Reynolds number for the particle Re_D , as per Cunningham's formula

$$C_D = \frac{24}{Re_D} \left(1 + \frac{3}{16} Re_D \right)^{1/2} \quad \text{if } Re \lesssim 100$$

$$Re_D = \frac{D_p U_s}{\nu_g}$$

The dynamic viscosity depends on ρ_g , specific mass of the gas and μ_g , absolute viscosity, and is found to be, from the composition obtained previously

$$\mu_g = 1.129 \times 10^{-4} \text{ m}^2/\text{s} \text{ (1.214 ft}^2/\text{s)}$$

U_s is then computed as a function of D_p from equation (A3-1), by iteration, and plotted in Fig. A3-1 (Case A). In particular,

Fig. A3-1. Settling Velocity vs. Particle Size for
Two Gas Compositions

for the smallest particle present in the mixture:

$$D_p = 0.104 \text{ mm}, U_s = 0.35 \text{ m/s (1.15 ft}^2/\text{s)}$$

The angle of repose of the material can be estimated at 45° . Radial lifters would start dropping the product when the radius to their point of attachment makes an angle of 45° with the horizontal, in the sense of rotation. The particle drop distance would be about $3/4$ of the dryer diameter, D_D .

For $D_D = 1.83 \text{ m (6 ft)}$, the fall time is of the order of $\frac{.75 \times 1.83}{.35} = 3.92 \text{ s}$. (As computed below, the exact time to reach 95% of V_s is 4.15 sec, since the settling velocity is not reached instantaneously.)

Entrainment effects are evaluated by determining the time T^* , needed for the particle to reach from rest a velocity V_p^* , under a horizontal force equal to the drag, for a relative speed, $V_g - V_p$:

$$M_p \frac{dv_p}{dt} = C_D S_p \rho_g \frac{1}{2} (V_g - V_p)^2$$

$$C_D = \frac{24}{Re_g} \left(1 + \frac{3}{16} Re_g\right)^{1/2} \quad Re_g \leq 100$$

$$Re_g = \frac{V_g - V_p}{\nu_g} D_p$$

or

$$\frac{dv_p}{dt} = \frac{3}{4} \frac{C_D}{D_p} \frac{\rho_g}{\rho_p} (V_g - V_p)^2 = F(V_p)$$

For given V_g , ν_g , D_p , C_D is a function of V_p only, and time T^* needed for the particle to reach velocity V_p^* ($< V_p$) is

$$T^* = \int_{V_p}^{V_p^*} \frac{dv_p}{F(V_p)} = G(V_p^*) \quad (A3-2)$$

Next, the time " t_w " needed for the particle to reach the wall in a vertical fall of $3/4 D_D$ is obtained from Gaudin (34)

$$s = \frac{U_s^2}{2g} \ln \left[\frac{1}{2} + \frac{1}{4} \exp\left(\frac{t^2 g}{U_s^2}\right) + \frac{1}{4} \exp\left(-\frac{t^2 g}{U_s^2}\right) \right] \quad (A3-3)$$

in which $s = 3/4 D_D$ and U_s is the settling velocity, as the independent variable. On Fig. A3-2, the dependence of t_w on V_s is plotted. It should be noted that the actual time needed to reach the wall will be larger than t_w , if there is any significant horizontal entrainment, so that using t_w from (A3-3) will result in conservative calculations in the following developments.

The horizontal displacement, x_w , over t_w , is obtained by numerical integration of (A1-2) in the form

$$v_p = G^{-1}(t) = \frac{dx_p}{dt}$$

$$x_w = x_p(t_w) = \int_0^{t_w} G^{-1}(t) dt \quad (A3-4)$$

For convenience, the time T_{95} needed to reach 95% of the gas velocity was computed from (A3-2), and the following (conservative) simplification was made,

$$\text{if } T_{95} \leq t_w: x_w = x(T_{95}^*) + 0.95 (t_w - T_{95}^*) V_g$$

Fig. A3-2. Time to Reach the Wall, t_w , vs. Settling Velocity

If $T_{95} > t_w$, $x_w = x(t_w)$ was determined by integral (A3-4). When needed, interpolation was used between two values of t , between which t_w was comprised.

For generality, x_w was plotted as the ratio of $\frac{x_w}{D_D}$, i.e., as a number of dryer diameters. In Fig. A3-3, Case A is referenced by the curve in full line. It is seen that, at the specified inlet conditions, particles 0.1 to 0.6 mm will be entrained horizontally over distances which are 1.8 to 0.11 dryer diameters, respectively.

The question to be addressed next is: what should be construed as "excessive entrainment distance?" For common values of the diameter 1.83 m (6 ft), angular speed (2 RPM) and residence times (20 minutes), the number of falls of a particle, in the absence of entrainment, is in the order of

$$\frac{1200}{30 \times \frac{3\pi}{4} / 2\pi} \approx 107$$

If the length to diameter ratio has the common value $L_D/D_D = 4$, then the axial progression should be approximately: $4/107 = 3.7\%$ of the diameter, per fall, on the average. For $\frac{L_D}{D_D} = 5$, it is 4.7% of the diameter, on the average.

If the entrainment distance between falls, x_w , is twice that distance, or approximately 7.5 to 9.5% of the diameter, this could be considered as a criterion for excessive entrainment of the particles.

In Fig. A1-3, it is seen that this criterion is met, for inlet conditions, for all particles smaller than 0.6 mm, or 91% of the feed.

Axial Displacement, Number of Dryer Diameters

Fig. A3-3. Axial Displacement (Dryer Diameters), Per Fall, as a Function of Particle Size

To be more precise, the calculations will now be carried out at a "downstream" location (Case B), so that "average" entrainment characteristics can be determined between Case A (inlet) and B (downstream).

Case "B": Downstream Condition

The thermodynamic conditions at the downstream location are obtained by assuming the temperature to be 149°C (300°F), and the gaseous mixture to be made of the original combustion gases and one-half of the total moisture to be carried away (240 kg/h for 4 Mg/h and 12% initial moisture)

$$p = 101,320 \text{ Pa}$$

$$r = 303.8 \text{ (J/kg, } ^\circ\text{K)}$$

$$\dot{M}_g = 0.789 \text{ kg/s}$$

$$\rho_g = 0.790 \text{ kg/m}^3 \text{ (0.0493 lb/cft)}$$

$$V_g = 0.446 \text{ m/s (1.462 ft/s) [15\% fill ratio, } D_D = 1.83 \text{ m (6 ft)]}$$

From the composition of the mixture of combustion gases and moisture, the dynamic viscosity is computed to be

$$\mu_g = 0.0219 \text{ (cp)}$$

Hence

$$\nu_g = 2.77 \times 10^{-5} \text{ m}^2/\text{s} \text{ (2.98} \times 10^{-4} \text{ ft}^2/\text{s)}$$

Settling velocity U_s .

The velocity U_s corresponding to the above gas conditions was computed as a function of the particle diameter D_p , and is plotted with the label "B" in Fig. A3-1. The curve giving t_w , time to reach the wall during a vertical fall from a height equal to 3/4 diameter, vs. U_s , as plotted in Fig. A3-2, still

holds but obviously the same abscissa (U_s) will not correspond to the same particle diameter for cases "A" and "B". The horizontal displacement x_w during a fall, expressed in dryer diameters, is computed as above, using the values of μ_g and V_g corresponding to case "B". The results are plotted in Fig. A3-3 (dotted line).

Again, it is seen that, under the conditions prevailing downstream, relatively close to the exit of the dryer, particles 0.1 mm to 0.6 mm in diameter will be entrained 0.58 to 0.026 diameters at each fall. For a length to diameter ratio L_D/D_D equal to 4, the entrainment distance will be at least twice the average design distance (3.7% diameter), for all particles smaller than 0.38 mm, or at least 37% of the feed product.

Average Conditions

If an arithmetic average between entrainment distances x_w (per fall) for cases A and B is taken as the criterion, Fig. A3-3 shows that, for a dryer length to diameter ratio, L/D , of 4, entrainment of particles will be excessive (in the sense defined above) for all particles smaller than 0.56 mm, i.e., at least 75% of the feed. For $L_D/D_D = 5$, this figure would be 60% and for $L_D/D_D = 6$, it would be 49%.

CONCLUSION

It is predicted that a very significant fraction (by weight) of the particles, 50% or more, will be entrained out of the dryer in a time significantly shorter than the average design dwell time. Whether the average moisture in the collected carryover fines is indeed negligible, without having to resort

to load recirculation, appears quite doubtful. This point could only be resolved by an extensive series of tests (which the manufacturers required, incidentally, to firm up their quotes).

In conclusion, dusting and carryovers of fines in direct drying represents a definite drawback in this application, dealing with a polydisperse mixture of fine particles in the 0.1 to 1.0 mm range, of relatively high initial moisture (12%).

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